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RECLAMATION OF ENERGY FROM ORGANIC WASTE

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RECLAMATION OF ENERGY FROM ORGANIC WASTE

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FOREWORD

Man and his environment must be protected from the adverse effects of pesticides, radiation, noise and other forms of pollution, and the unwise management of solid waste. Efforts to protect the environment require a focus that recognizes the interplay between the components of our physical environment--air, water, and land. The National Environmental Research Centers provide this multidisciplinary focus through programs engaged in

- studies on the effects of environmental contaminants on man and the biosphere, and
- a search for ways to prevent contamination and to recycle valuable resources.

In an attempt to solve the problems involved in solid waste disposal, this study applied an anaerobic fermentation process to the production of methane from an organic fraction of urban refuse.

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ABSTRACT

The continued pressure for a clean burning fuel has created such a demand for natural gas that the gas reserves in this country are being severely depleted. The mounting quantities of urban solid wastes presents a significant source of supplemental energy if a practical means of extracting the energy can be developed. A process employing microorganisms capable of fermenting organic compounds to methane and carbon dioxide has been used for decades to stabilize the organic sludges generated by water pollution control processes. The biological mechanism has been extensively studied, elucidating the nutritional, environmental and operational requirements.

This study applied the anaerobic fermentation process to the production of methane from the organic fraction of urban refuse. Shredded domestic refuse from which the inorganic fraction was separated was used as a substrate. Raw sewage sludge was added to the substrate in proportion to the rate at which it is produced by a population producing a given quantity of refuse. The quantity and quality of gas produced, the rate of gas production, the solids reduction, nutritional requirements and operating problems were evaluated in a laboratory system operating at temperatures ranging from 35°C to 60°C.

The results of the laboratory study together with published data on both capital and operating costs of refuse shredding, refuse separation, reactor volume, reactor mixing, reactor heating and residue dewatering were used to analyze the economics of the process. This analysis indicated that methane can be produced by anaerobic fermentation of organic refuse at a cost that would permit the sale of the gas at a price that is competitive with the current energy costs.

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CONCLUSIONS

An effective microbial system was developed for producing methane from organic refuse. As long as due consideration was given to the nutritional and environmental requirements of the micro-organisms, the system functioned very well.

The addition of raw sewage sludge to the reactors markedly improved the biological conversion even though adequate major nutrients, nitrogen and phosphorus, were supplied separately.

Inadequate alkalinity was present in the reactors to maintain the pH above 6.6. It was necessary to add caustic for pH control. The caustic requirements decreased with increasing temperature and retention time.

At 35°C, the gas production was 1.4 and 3.05 scf per pound of refuse added at 4 and 30 days retention time respectively. At 60°C, the gas production was 4.23 and 4.96 scf per pound of refuse added at 4 and 30 days retention time respectively.

The methane content of the dry gas produced varied from 54 to 70 percent. The methane content decreased with increasing retention time and increasing temperature.

A temperature of 42.5°C yielded the maximum gas production in the mesophilic temperature range while the maximum gas production in the thermophilic temperature range occurred at 60°C.

An economic analysis of the unit processes, excluding shredding, separation and disposal costs, for a potential system processing 100 tons per day of dry refuse yielded the following conclusions:

1. For all combinations of operating conditions, amortization of capital accounted for a major portion of the total annual costs. For feed slurry concentration of 20 percent, capital amortization accounted for 44 to 69 percent of the total costs.
2. At high reactor slurry concentrations and long retention times, mixing costs account for approximately one-fifth of the total annual costs.
3. The heating costs at the 60°C temperature accounted for 39.3 percent of the total annual cost for a high feed slurry concentration and short retention time. The heating costs at 35°C accounted for less than 10 percent of the total annual costs.

4. The maximum net benefits in the mesophilic range occurred at the higher feed slurry concentrations, but net benefits were relatively insensitive to reactor retention times.
5. The maximum net benefits in the thermophilic range occurred at the maximum feed slurry concentration and the minimum retention time tested.
6. The cost of producing gas by this process was 11.0 and 7.8 cents per 1000 scf at the optimum mesophilic and thermophilic temperature respectively with a feed slurry concentration of 20 percent solids.

The shredding and separation processes substantially increase the cost of producing gas. Using a cost for these processes of \$2.00 per ton, the production cost of the gas is increased by 27 and 24 cents per 1000 scf for the optimum mesophilic and thermophilic temperature respectively.

A benefit will accrue from the disposal of the residue. Processing 100 tons of refuse as received will yield 73.4 to 89.4 tons of residue that will require only one-third the landfill volume required of the unprocessed refuse.

RECOMMENDATIONS

The requirements for an effective biological conversion of the refuse to methane have been elucidated. The major work that needs to be undertaken involves an evaluation of the physical processes accompanying the biological process. Therefore, it is recommended that a large scale pilot study be undertaken to evaluate the physical processes. This study should develop the necessary design parameters for the shredding and separation of the raw refuse, mixing and heating requirements for the reactor, residue dewatering and disposal, and gas purification. A simulation model should be developed to predict the operation of the system under various operating conditions such that an optimum or near optimum design of a full scale system can be obtained.

Upon completion of an economic evaluation of the pilot study, a decision can be made regarding the economic attractiveness of the process. Based upon the economic analysis conducted in this study, the process has potential for profitable operation. These analysis should be verified and/or adjusted with a pilot scale study.

INTRODUCTION

Among the many challenges confronting the world as a whole, and this country in particular, are two seemingly unrelated problems. The first and perhaps the more obvious of the two is a problem of solid waste management. In particular, it is a problem of solid waste disposal. The second is the decline in energy resources as evidenced by the increasing shortage in natural gas reserves. The development of a process to convert organic refuse into methane would provide at least a partial solution to both of these problems.

Solid Waste Disposal

In solid waste management programs disposal always plays a major role. This is a result of the ever-increasing amounts of solid waste generated in this country. On a per capita per day basis it was estimated that the urban refuse contribution alone amounted to seven pounds in 1967 (Eliassen *et al.*, 1969). The figure increases to ten pounds when industrial solid wastes are included. The problem is indeed monumental, particularly when one considers the increased urbanization of the population of the United States.

Existing techniques for the disposal and processing of solid wastes consist primarily of landfill and incineration. Composting is used on a very limited basis. In a broad sense all current disposal practices are concerned with the return of the waste material to the land in either of two forms. In the raw form solid wastes are disposed of in landfills. In the processed form, incinerator residue must also find its final repository in landfills. Where composting is practiced, it too is returned to the land as a soil additive.

Metropolitan areas are finding the disposal problem ever more difficult as nearby landfill sites become exhausted. Composting suffers the disadvantages of high production costs as well as a very limited market for the final product. As a result it has fallen into disfavor and few communities practice it any longer. Therefore, with present technology the landfill must be the ultimate recipient of the generated solid wastes. As the land areas available for use as landfill sites become more and more distant from the source of solid waste generation, the costs attributed to the transportation of these wastes increase. It follows that volume and weight considerations become very important. Incineration currently provides reductions in the quantity of refuse up to 70 or 80 percent depending upon the refuse composition. This process is expensive and is likely to become even more so as air pollution standards become more stringent. A process achieving reductions in weight and volume at economics competitive with incineration, and not suffering its air pollution control limitations, would be very attractive indeed.

Energy Shortage

The total energy demand in the United States rose from 45 quadrillion BTU in 1960 to over 60 quadrillion BTU in 1970 and is expected to top the 90 quadrillion BTU mark by 1980. By the year 2000, this energy demand is forecasted to reach 200 quadrillion BTU (*Chemical and Engineering News*, 1970a). The share of natural gas (methane) in this energy market was about 31 percent in 1968 (*Chemical and Engineering News*, 1970b). This total energy demand, and in particular that on natural gas, cannot continue without exhausting the available fuels in the not unforeseeable future. With the exception of natural gas, natural resources are currently abundant but not unlimited and efforts should be made to conserve these resources.

The outlook for continued expansion of natural gas consumption is not particularly good. Gains in production of gas between 1947 and 1968 averaged more than six percent a year with the production growing from 5.6 to 19.3 trillion cubic feet. During the same period reserves increased on an average of 2.5 percent annually or from 165 to 282 trillion cubic feet. The reserves-to-product ratio has declined from 29.5 to 14.6 years (*Chemical and Engineering News*, 1970b). By the end of 1969 the reserves were down to 275.1 trillion cubic feet while production rose seven percent to 20.7 trillion cubic feet (*Chemical and Engineering News*, 1970c). Even with improvements to reserves it is estimated that production of gas in the 48 contiguous states will be limited to about 27 trillion cubic feet a year for 1974 to 1990. With gas from Alaska, U. S. production could reach more than 35 trillion cubic feet a year by 1990 and hold at this level for nearly ten more years (*Chemical and Engineering News*, 1970b). Figure 1 illustrates the forecasted future situation in natural gas requirements and supplies.

Even with optimistic views on reserves and future production, demand for gas will remain greater than additions to reserves. An increase in additions to reserves will only give lead time to develop other sources of gas--imports from Canada, imported liquid natural gas and coal gasification (*Chemical and Engineering News*, 1970b).

Refuse as a Source of Fuel Gas

In wastewater treatment anaerobic digestion has been used for a long time to effectively reduce the quantity of organic sludges and to transform them into stable, easily dewaterable residues. The reduction is accomplished by the biological conversion of the organic solids to methane and carbon dioxide. This anaerobic process has a very low efficiency of biological energy conversion.

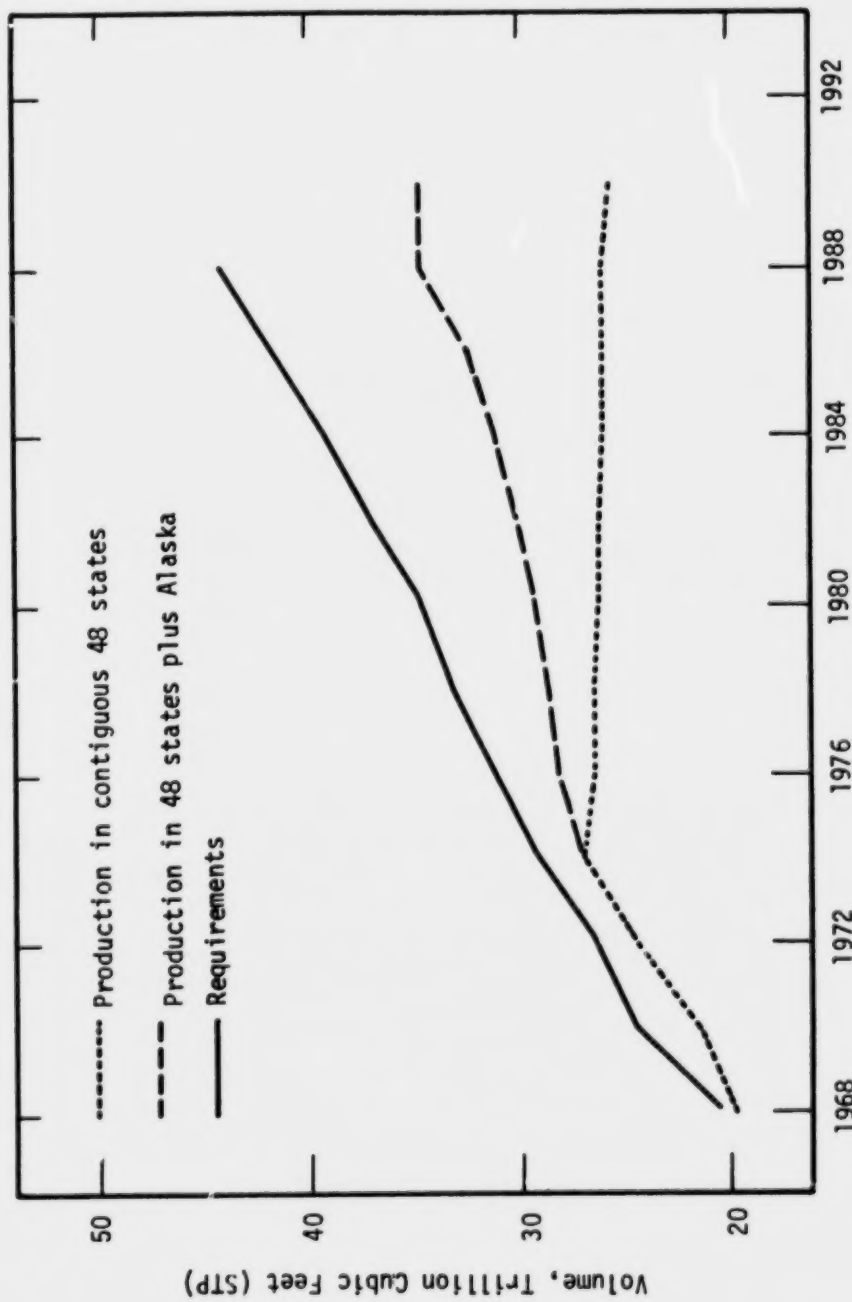


Figure 1. Future Natural Gas Supplies and Requirements
(*Chemical and Engineering News*, 1970b)

Most of the energy of the substrate is lost. The gas produced has a high calorific content and is a valuable end product with potential for reclamation. The heat value of the gas produced from the digestion of sewage sludges ranges from 7000 to 9000 BTU per pound of organic matter destroyed, depending on the nature and composition of the solids to be digested.

Various studies have demonstrated that organic refuse, particularly garbage, is amenable to anaerobic decomposition as long as the proper environmental conditions of temperature, pH, absence of oxygen, etc. and adequate nutrients are maintained. These will be reviewed in more detail later. Organic refuse is composed basically of the same compounds, carbohydrates, proteins and fats, as sewage sludge though in different relative proportions. There appears to be no reason why refuse cannot undergo adequate anaerobic decomposition. Proper environmental conditions must be maintained, the deficient nutrients supplied and toxicity problems eliminated.

In addition to providing a means of energy conservation this process provides a significant reduction in the quantity of waste material for ultimate disposal. Assuming that about 70 percent of domestic refuse is biodegradable and that 70 percent of the biodegradable portion is gasified in an anaerobic process, the resulting overall reduction in weight would amount to about 50 percent. This will greatly reduce transportation costs to the landfill sites as well as extend the life of the sites. Inherent in this is the assumption that the solids in the digester effluent will dewater easily.

Moreover, this residue exhibits a very important advantage over the present solid wastes that are placed in landfills. Untreated refuse has not been stabilized biologically and is in a process of decay. When placed in a landfill it decomposes and can pollute the groundwater if adequate preventive measures are not taken. It can also cause substantial settlements in the fill area that may render it unsuitable for any kind of structural support and real estate development. These problems are far less severe with digester residue since it has been biologically stabilized. The development of systems of this type will be a big contribution to the overall problem of solid waste management.

Assuming the degree of conversion mentioned above, a per capita contribution of domestic refuse of seven pounds per day and a heat value for the refuse digester gas equal to that of sewage digester gas, the per capita contribution of energy would range between 24,500 BTU and 31,500 BTU daily. With an average value of 28,000 BTU per capita per day, this translates to about 2.0 quadrillion BTU per annum for the whole population of the United States, about 3.3 percent of the total 1970 United States energy

demand. This will satisfy about 11 percent of the 1970 demand for natural gas. In terms of dollars, using a fuel value of 50 cents per million BTU, this means a fuel value of about one billion dollars per year. This, of course, is a gross figure and does not include the cost of the refuse conversion. In addition this gas would be produced in the urban areas, eliminating the need for long distance transportation.

Purpose of the Study

This study was designed to evaluate the potential for reclaiming methane gas from organic refuse by means of anaerobic fermentation. The mechanism of methane fermentation has been studied extensively, elucidating the constraints that must be satisfied for the system to function effectively. The problems likely to be encountered in the fermentation of organic refuse were unknown. It was the purpose of this laboratory study to define these problems and develop information to evaluate the feasibility of using this process on a large scale.

Studies were undertaken to determine the response of the methane fermentation process to external variables such as temperature, retention time, pH and feed slurry concentration. Potential operational problems were identified. The environmental and nutritional requirements for the fermentation of urban refuse were evaluated. The quality and quantity of gas production were determined for various operating conditions. From the data collected it was possible to evaluate the practicality of employing this process as a means of reducing the quantity of refuse for disposal while reclaiming a useful product, methane.

THEORETICAL CONSIDERATIONS

Basic Fundamentals of Anaerobic Digestion

Anaerobic treatment of complex organic materials is considered to be basically a two-stage process as indicated in Figure 2. In the first stage, a group of facultative and anaerobic bacteria, known as the "acid formers," act upon the complex organics and change the form of complex fats, proteins and carbohydrates to simple soluble organic materials. The end products of this first stage conversion are primarily short chain organic acids, also known as volatile acids, and small amounts of bacterial cells. This stage accomplishes no stabilization of the waste material but it is an essential prerequisite for the second stage in which the actual stabilization of the waste matter occurs. It places the organic matter in a form suitable for the second stage of treatment.

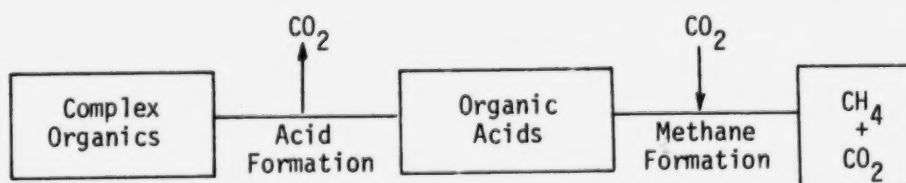


Figure 2. Anaerobic Stabilization of Complex Organics

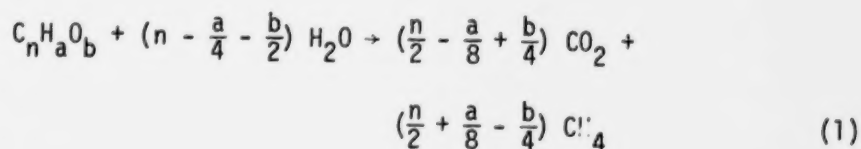
In the second stage, the short chain organic acids are acted upon by a strictly anaerobic group of bacteria, termed the "methane formers," and are converted to gaseous end products, methane and carbon dioxide. The methane formed in this second stage, being insoluble in water, escapes from the system. It can be collected for use as a fuel. The carbon dioxide involved partially escapes in the gaseous form and partially goes into solution. It is in this second stage that stabilization occurs through the removal of oxygen demanding material in the form of methane gas. Cell production is also minimal compared to aerobic processes. This is a direct result of the high energy content of the products, in particular methane (McKinney and Conway, 1957). This is an advantage as the amount of solids requiring ultimate disposal is minimized by eliminating any significant microbial protoplasm production in the process of stabilizing the organic material.

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There are several different groups of methane formers with each group characterized by its ability to ferment only a specific number of compounds. Therefore, several different bacteria are required for stabilization of the organic material (McCarty, 1964a). The most important of these methane formers, those utilizing acetic and propionic acids for substrates, grow quite slowly and for most organic compounds are rate limiting at the lower retention times. For sewage sludge digestion, Pfeffer (1968) found that in systems operating with solids retention times of approximately ten to fifteen days, the rate limiting step is methane fermentation. When such systems are operated at solids retention times greater than fifteen days, the rate limiting step then becomes the hydrolysis of organic solids.

Tracer studies have indicated the major sources of methane (McCarty, 1964a). One source is the direct cleavage of acetic acid into methane and carbon dioxide. The methyl carbon of acetic acid, together with its three hydrogen atoms, is converted intact into methane gas. The carbonyl carbon is converted to carbon dioxide. Most of the remaining methane is formed from the reduction of carbon dioxide. Carbon dioxide, functioning as an electron acceptor, is reduced by hydrogen atoms enzymatically removed from organic compounds. The availability of carbon dioxide in such a step is never a rate limiting factor as there is always a large excess of it from the bicarbonate buffer system that is present in anaerobic systems (McCarty, 1964a).

Organic destruction in anaerobic treatment is directly related to methane production and vice versa. Buswell and Mueller (1952) developed Equation 1 to predict the quantity of methane from a knowledge of the chemical composition of the waste:



McCarty (1964a) showed that the theoretical methane production from the complete stabilization of one pound of BOD_L or COD was 5.62 cubic feet at standard temperature and pressure. From this he forwarded Equation 2 for estimating methane production from waste strength:

$$C = 5.62 (eF - 1.42A) \quad (2)$$

where: C = cubic feet of CH₄ produced per day (STP)
 e = efficiency of waste utilization
 F = pounds of BOD_L added per day
 A = pounds volatile biological solids produced per day.

The efficiency of waste utilization, e, should not be confused with the stabilization efficiency. It is a factor designating the efficiency of the conversion of the organic waste to both the gaseous form and to biological solids. Whereas the former form represents waste stabilization, the latter conversion represents no such thing. The pounds of volatile biological solids produced per day, A, can be estimated from Equation 3 (McCarty, 1964d):

$$A = \frac{aF}{1 + b(\text{SRT})} \quad (3)$$

where: a = growth constant
 b = endogenous respiration rate
 SRT = solids retention time in days.

Values for a and b are shown in Table 1 for various organic compounds (Speece and McCarty, 1964).

Table 1. Growth Constants and Endogenous Respiration Rates

Substrate	Growth Constant a	Endogenous Respiration Rate, b
Fatty Acid	0.054	0.038
Carbohydrate	0.240	0.033
Protein	0.076	0.014

The percentage of added BOD_L which is stabilized, S, is given by Equation 4 (McCarty, 1964d):

$$S = \frac{100C}{5.62 F}$$

$$= \frac{100(eF - 142 A)}{F} \quad (4)$$

The efficiency of waste stabilization is related to the solids retention time. As the solids retention time decreases the relative proportion of active cells washed out of the system increases. If the solids retention time falls below a certain limit, the microorganisms responsible for anaerobic digestion will be washed out faster than they can reproduce themselves and the result is failure of the process. The minimum retention time is dependent on the temperature and the types of microorganisms in the system under consideration. This latter is dependent on many factors, one of which is the type of substrate being utilized. Though it is possible to operate near the minimum retention times, efficiencies are low and process dependability is poor. Solids retention times of at least two and a half times the minimum are recommended (McCarty, 1964d).

While there are many different methane forming bacteria, there are also many different acid forming bacteria. For attaining good efficiencies of waste stabilization, a proper balance among all these different organisms is required. The establishment and maintenance of this balance dictates operation under optimum environmental conditions. These are discussed in brief in the next paragraphs.

Temperature is a very important operational parameter in anaerobic fermentation process. As temperatures increase, rates of reaction proceed much faster and this results in more efficient operation and lower retention time requirements (McCarty, 1964b). Two optimum temperature levels have been established. In the mesophilic level the temperatures range from 30°C to 37.5°C and in the thermophilic level, they range from 49°C to 51°C. Although the rates of reaction in the thermophilic level are much faster than those in the mesophilic level, the economics of most sewage sludge digestion systems have dictated operation in the mesophilic level or lower (McCarty, 1964b). This stems from the fact that the methane requirements to maintain thermophilic temperatures in most digesters are excessive and uneconomical. This, in turn, is a result of the inability to thicken the feed sludges sufficiently such that the organic loading and the resulting methane production per unit digester volume are sufficiently high to make such an operation economically attractive.

Another environmental requirement for anaerobic treatment is the maintenance of anaerobic conditions in the digester. The methane formers are strict anaerobes and even small amounts of oxygen can be quite detrimental to them. This necessitates, in most cases, a closed digestion tank which is also convenient because collection of the methane produced is facilitated.

The third environmental requirement for optimum operation is that for a proper pH. McCarty (1964b) reports that anaerobic treatment can proceed quite well with a pH varying from about 6.6 to 7.6 with an optimum range of about 7.0 to 7.2. Beyond these limits anaerobic digestion proceeds with decreasing efficiency until at a pH of 6.2 and lower, the acid conditions become quite toxic to the methane formers and waste stabilization comes to a virtual halt.

Control of pH should be exercised when the pH appears likely to drop below a value of 6.6. This is done by the addition of an alkali. In sewage sludge digestion the use of lime for such control has been the most widespread, but because of its many advantages over lime, sodium bicarbonate has lately been receiving increasing attention as a substitute for lime for pH control.

The bacteria responsible for waste conversion and stabilization in the anaerobic process require nitrogen, phosphorus and other materials in trace quantities for optimum growth. Therefore, another important environmental condition is the presence of the required nutrients in adequate quantities. Municipal waste sludge usually contains all the required nutrients in adequate quantities but other substrates, industrial and solid wastes in particular, might not. If the nutrients are not present in the required quantities, they must be either added or supplemented.

McCarty (1964a) calculated the nitrogen and phosphorus requirements based on an average chemical formulation of biological cells of $C_5H_9O_3N$. This yielded a nitrogen requirement of about 11 percent of the cell volatile solids weight and a requirement for phosphorus equal to about one-fifth of this figure. Although these requirements should theoretically be based on the fraction of waste removed during treatment rather than on the waste added, it is good practice to base them on waste additions (McCarty, 1964a). Other elements having stimulatory effects at low concentrations include, but are not limited to, sodium, potassium, calcium, magnesium and iron (McCarty, 1964c). All of these preceding elements exhibit inhibitory effects at higher concentrations.

The fifth and final environmental requirement for successful anaerobic treatment is that the waste be free from toxic materials. This is particularly true for concentrated organic wastes which, though normally are more susceptible to anaerobic treatment, are also more likely to have high or inhibitory concentrations of various materials ranging from inorganic salts to toxic organic compounds.

Some alkali and alkaline earth-metal salts above certain concentrations exhibit degrees of inhibition and toxicity. The threshold levels vary depending on whether these metals act singly or in

combination. Certain combinations have synergistic effects, whereas others display antagonistic effects.

Ammonia, particularly when in the ammonia form, is inhibitory when present in high enough concentrations. At concentrations between 1,500 and 3,000 mg/l and a pH greater than 7.4, the ammonia concentration can become inhibitory. At concentrations above 3,000 mg/l, the ammonium ion itself becomes quite toxic regardless of pH (McCarty, 1964c).

Other common forms of toxicity include those of sulfides, heavy metals and toxic organic materials. Concentrations of soluble sulfide varying from 50 to 100 mg/l can be tolerated in anaerobic treatment with little or no acclimation, whereas concentrations up to 200 mg/l can be tolerated with some acclimation (McCarty, 1964c). Low soluble concentrations of copper, zinc and nickel salts are associated with most of the problems of heavy metal toxicity in anaerobic treatment. Also, there are many organic materials that exhibit inhibitory effects. These range from organic solvents to many common materials such as alcohols and long chain fatty acids in high concentrations (McCarty, 1964c).

It is well to know that microorganisms usually have the ability to acclimate to some extent to inhibitory concentrations of most materials if the process is acclimated to the inhibitory substance. Also, it should be recognized that only materials in solution can be toxic to biological life (McCarty, 1964c). Control of toxicity or inhibition can, in general, be achieved by one or more of three ways: 1) the removal from the waste stream or inactivation of toxic materials by such means as chemical precipitation, 2) the dilution of the waste stream below the "toxic threshold" of the toxicity causing material, and 3) the addition of an antagonistic material.

Anaerobic Digestion of Domestic Refuse

As mentioned earlier, the solids retention time is one of the most important factors determining the degree of waste stabilization. Increasing solids retention times result in increasing stabilization efficiencies and higher gas production. Figure 3 (Rankin, 1948) illustrates the relationship between solids retention time and the degree of stabilization achieved for sewage sludge. However, especially where the waste is dilute, high retention times become associated with excessively large digestion tanks and also with excessively high thermal requirements for heating. At a certain minimum concentration, the anaerobic process, as conventionally known, ceases to be economically feasible and either the anaerobic contact process or a different method of treatment is employed. The anaerobic contact process is based on exactly the same principles as its conventional counterpart, except that solids

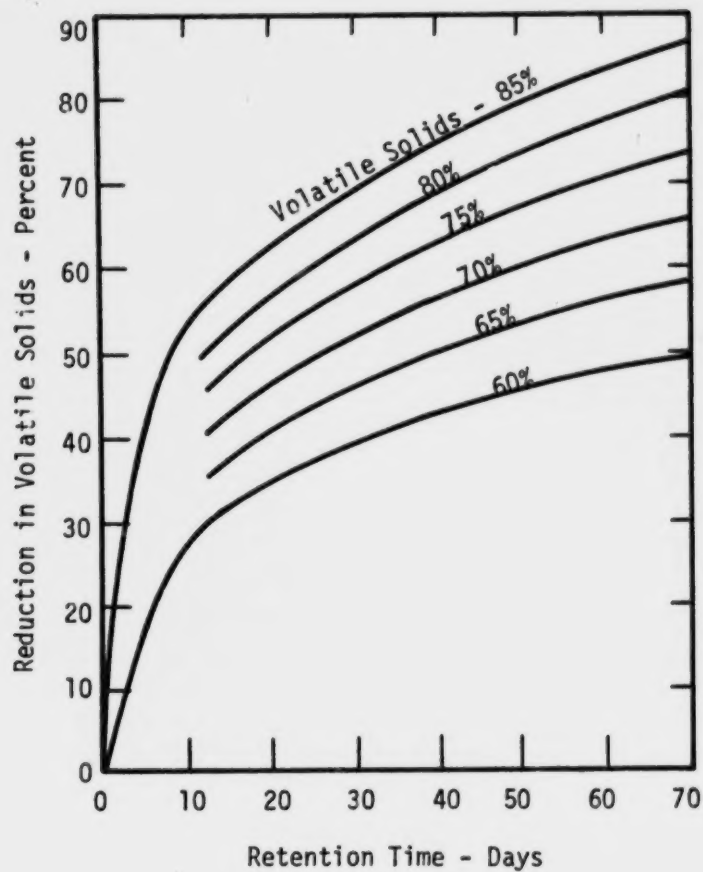
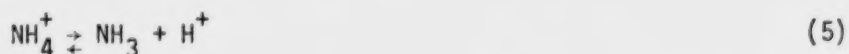


Figure 3. Variation of Volatile Solids Destruction with Detention Time (Rankin, 1948)

return is practiced. This way, the bacteria are not lost with the effluent but are maintained in the system. This also permits the maintenance of high solids retention times at low liquid retention times.

No such special arrangements need be made when digesting organic solid wastes. The moisture content of the feed slurry can be adjusted to the required level as it only involves the addition of water in the desired amounts. Therefore, this process will capitalize on one advantage refuse has over sewage sludge, i.e. a high solids content. It will allow for high loadings on the systems while still maintaining the solids retention times so necessary for stabilization of the organic material. With substrates rich in proteins, concentrated feed slurries pose problems. As the concentration of the solids in the feed slurry increases, so does the ammonia concentration and the alkalinity. If the volatile acids remain low due to effective methane fermentation, the high alkalinities will cause an increase in the pH of the system. This rise in pH causes a shift to the right in the following equilibrium equation:



If this happens, inhibition becomes related to the ammonia concentration. As ammonia is inhibitory at a much lower concentration than the ammonium ion (McCarty, 1964c), process failure could result as explained in a previous section. This, however, is not expected to be a problem with domestic refuse. With garbage assuming an ever-decreasing role in the composition of domestic refuse, it would be indeed fortunate if enough proteins were present to even partially satisfy the nutrient requirements for nitrogen. Golueke and McGauhey (1967) estimated that about 67 percent of the refuse that can be converted biologically was found to be paper. Garbage accounted for only 12 percent of the biodegradable refuse. The total nitrogen content of urban refuse is extremely small unless some industrial solid waste is contributing nitrogen. With this proportion of carbohydrates and proteins in the feed, the system is not expected to operate at high pH values or be faced with an ammonia toxicity problem.

Concern has been expressed over the rate of biological breakdown of such materials as celluloses, hemicelluloses and lignins. With the exception of lignins which are known to be very resistant to biological decomposition, it is expected that this will not pose any serious problems. The percentage of lignineous materials in typical domestic refuse, it is felt, will be small and restricted to certain types of wood and newsprint (Golueke, 1970).

Cellulose Reduction in Anaerobic Digestion

That cellulose hydrolysis could be the rate limiting step in digestion of wastewater sludges was postulated by Maki (1954) in his study of pure and mixed cultures of cellulolytic bacteria removed from a municipal digester. The characteristics of the undigested sludge which he used (see Table 2) is similar to that of organic refuse. Therefore, it might be expected that the hydrolysis of cellulose is the overall rate limiting step. This assumption was based on the fact that acid production from the hydrolysis and fermentation of cellulose was slower than the conversion of these products to methane and other end products.

Table 2. Chemical Components of a Sewage Sludge

Component	% Dry Weight
Hemicellulose	6
Cellulose	34.5
Lipids	14
Protein	19
Ash	34

Source: Maki (1954)

Kinetic studies of the anaerobic digestion process by Lawrence and McCarty (1967) and Andrews *et al.* (1964) stated that methane production was the overall rate limiting step. Pfeffer (1968) pointed out that at solid retention times greater than ten to fifteen days, hydrolysis of the complex organics becomes the rate limiting step. As domestic refuse has as high or higher concentrations of complex organics as the typical sludge on which these conclusions were based, it is logical to assume that hydrolysis of the complex organic materials will be the rate limiting step at shorter solids retention times.

A recent study by Chan and Pearson (1970) on the hydrolysis of cellulose indicated that hydrolysis of insoluble cellulose to soluble cellobiose appeared to be the rate limiting step in the anaerobic decomposition of cellulose. They showed that this rate could be accurately characterized by the Michaelis-Menten kinetic model and gave the different kinetic constants and coefficients that apply to the fermentation of cellulose. They also concluded that not only is the cellulose hydrolysis to cellobiose the limiting step

of the cellulose hydrolysis process, but also is the rate limiting step in the overall cellulose fermentation process.

The structure of cellulose, shown in Figure 4, indicates that the basic unit is the disaccharide cellobiose. This structure shows a recurrency period of 10.25 angstroms which is double that of a glucose unit using Haworth's pyranose ring structure.

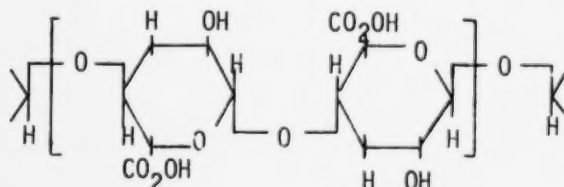
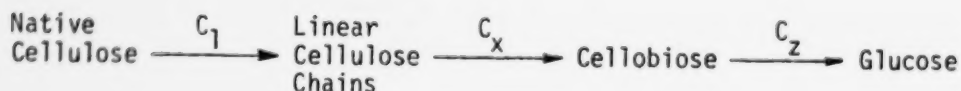


Figure 4. Structure of Beta 1,4 Cellulose

Cellulose exists in nature as the fibrous material constituting the cell wall of plants. The cellulose molecule varies in size with its source. The degree of polymerization, number of glucose units within a molecule, varies from 700 for cellulose from wood to nearly 4000 from cotton. The average molecular weight of cellulose is 400,000 corresponding to 2500 glucose units. Cellulose is not a homogenous chemical compound. Celluloses are all built up by the same substances but they may differ in average chain length, distribution of chain lengths and macromolecular structure. Additionally, the types of non-cellulosic substances contained within the cellulose structure may differ.

Reese and Levinson (1952), while studying the enzymatic hydrolysis of cellulose, proposed the following pathway for cellulose decomposition.



Little was known about the mechanism of C_1 . More recent works by Reese and others indicate that there is still little quantitative knowledge about this step C_1 . Its presence is assumed due to the lack of ability of certain organisms to utilize native cellulose yet being able to attack linear cellulose. C_x represents various cellulases. These enzymes break linear cellulose down primarily to cellobiose. Other compounds have been reported to have been produced by cellulases' hydrolysis of linear cellulose. These

other products include cellotriose, cellotetrose, glucose and disaccharides other than cellobiose. The mode of hydrolysis, as described by Reese (1956), is assumed to be both random attack of linkage and removal of end units.

Tests run by Reese *et al.* (1950) indicated the effect of pH on various cellulases. Tests were performed on filtrates from various cellulolytic bacteria indicating highest activity at pH values between 5.0 to 7.0. Similar tests to show temperature effects indicated continual increase in cellulase activity up to 55°C.

Bacterial Decomposition of Cellulose

Mesophilic Bacteria

Von Tappeiner (1884) first showed the importance of mesophilic bacteria on hydrolysis of cellulose. His works were associated with cellulolytic bacteria in the cattle rumen. Omelianski is reported to be the first to isolate a cellulolytic bacteria in 1902. His studies of by-products of cellulose decomposition indicated the release of methane and carbon dioxide from one organism and only hydrogen and carbon dioxide from a second organism. Subsequent observers have not reported by-products similar to Omelianski's second pure culture.

Pringsheim (1917) first found cellobiose and glucose to be intermediates in the decomposition of cellulose. Enebo (1949) found the same pathway for the thermophilic bacteria.

Heukelekian (1927) studied the effect of pH on cellulose digestion by measuring the decomposition rate in limed and unlimed digesters. Without liming the pH fell from 5.8 to 4.2 in seven days and then gradually increased. Buffering with lime to a pH of 7.4 yielded 73 percent cellulose destruction within seven days. This compares to 30 percent removal in the unlimed digester. Heukelekian concluded that lime stimulates cellulose decomposition.

Another study by Heukelekian (1928) showed the effect of cellulose content on gas production. He also studied the effect of lignin, hemicellulose and complex celluloses on digestion. Digestion of high grade, pure cellulose toilet tissue gave gas production of 0.57 liters of gas/gm of cellulose. Lower gas production with higher percentages of carbon dioxide was found for lower grade toilet tissue and newspaper. A noticeable lag time for initiation of gas production caused Heukelekian to conclude that establishment of proper flora to be the rate limiting step in cellulose decomposition.

Hungate (1950) isolated mesophilic bacteria from several different sources. Optimum temperature for bacteria isolated from the rumen

of cattle was found to be 38°C to 40°C. Isolation of three strains from sewage sludge classified the bacteria as gram negative rods and diploids, extremely variable in size, particularly when growing on cellulose. Optimum temperature appeared to be 38°C. Sampling from different stages in the two stage anaerobic digestion process found no cellulolytic bacteria in the raw sludge. Cell counts by the dilution method indicated that cellulolytic bacteria concentration increased from the influent of the first stage to the effluent in the second.

Maki (1954) isolated ten strains of cellulolytic bacteria from the Pullman, Washington wastewater plant digester. Cellulose accounted for over 50 percent of the solids in the digester. The digester varied in pH range of 6.2 to 7.8 with 6.8 being average. Cell counts on the digester sludge indicated up to 10^6 organisms per liter. Studies on the pure cultures indicated three strains to be highly cellulolytic. Comparison of cellulose removal by mixed culture yielded a rate two and a half times greater than that of the most cellulolytic pure culture. This implies that there is synergistic action of by-product removal of the wastes produced by cellulolytic bacteria by non-cellulolytic bacteria. Maki suggests that the rate of fermentation of cellulose to be the rate limiting step in wastewater sludge digestion.

Dubos (1938) studied the effect of pH and availability of nitrogen on the density of cellulose decomposing bacteria in soil. Table 3 shows the optimum pH to be 6.5 to 8.2 by cell count. However, pH had little effect on cellulolytic bacteria density when nitrogen as ammonium sulfate was added.

Table 3. pH Effect on Density of Cellulolytic Bacteria

pH of Soil	Number Cellulolytic Bacteria (Bacteria/gm)	pH of Soil	Number Cellulolytic Bacteria (Bacteria/gm)
8.7	0	6.5	2.5×10^7
8.5	2.5×10^5	6.0	2.5×10^5
8.2	2.5×10^7	5.2	0
7.5	2.5×10^7	4.5	0
7.0	2.5×10^7	4.5	0

Thermophilic Cellulolytic Bacteria

McBee (1948) summarized characteristics of thermophilic cellulolytic bacteria. Morphologically, he described them as "slender, often slightly curved, gram negative, spore-forming rods." All pure cultures are obligate anaerobes. Optimum temperature range is 55°C to 65°C. Growth occurs in pH range of 6.4 to 7.4. Growth and fermentation occurred on cellulose, cellobiose, xylose and hemicellulose in tests on pure cultures.

Enebo and Pehrson (1960) studied the thermophilic digestion of cellulose from a pulping factory wastewater seeded with domestic sludge. Rapid breakdown of cellulose was reported at a ratio of cellulosic waste to domestic sludge of three to one. At a one to one ratio, 85 percent of the cellulose was removed in only eight days. This was achieved at a two percent initial solids concentration. The cellulose concentration decreased from 9900 mg/l to 1500 mg/l. Optimum temperature was between 55°C and 60°C. A significant feature of the pulp factory waste was the high lignin content. The cellulose measurement procedure was not adequately explained.

Chan and Pearson (1970) summarized the reported hydrolysis rates shown in Table 4. Comparison of these rates shows that there may be only little significance in increase of hydrolysis due to thermophilic digestion. However, consideration must be given for differences in systems, cultures and substrates utilized in these uncoordinated studies. Synergistic relationships at the microorganism level are important as shown by Stranks (1956). His method of by-product removal was dialysis. In mixed culture systems this by-product removal would be done by other organisms. Wastewater sludges have been shown to contain significant quantities of cellulose. Therefore, the known higher rate of digestion at thermophilic temperatures would suggest higher cellulose destruction rates.

Gas Composition

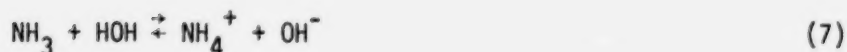
In the conversion of carbohydrates to carbon dioxide and methane, equal volumes of each gas are produced. This reaction is shown in Equation 6. However, all of the carbon dioxide is not released as a



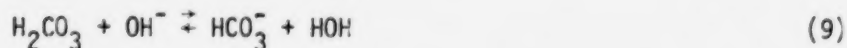
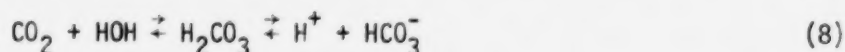
gas but enters into reactions with water and hydrogen ion. Microorganisms can deaminate biodegradable protein producing ammonia which reacts with water according to Equation 7. The hydroxide produced by this reaction reacts with carbon dioxide to form

Table 4. Summary of Hydrolysis Rate of Cellulose in Anaerobic Fermentation

Authority	System and Culture	Initial Cellulose Concentration (mg/l)	Cellulose Material	pH	Hydrolysis Rate (mg/l-day)
Maki	Batch, mixed 2 pure cultures from sewage, 38°C, mesophilic	2,000	Whatman's #1 filter paper	6.8	(1) 260
					(2) 660
Heukelekian	Batch, mixed culture from sewage, 25°C, mesophilic	3,120	Cellulose in sewage sludge	7.4	142
McBee	Batch, pure culture from soil and manure, 55°C, thermophilic	(1) 744	Absorbent cotton	-	(1) 149
		(2) 2,980		-	(2) 426
Stranks	Batch, mixed culture from rumen, 60°C, thermophilic	41,200	Whatman's #2 filter paper	6.5	11,400



bicarbonate ion as shown in Equations 8 and 9. Therefore, the protein content of the substrate will significantly effect the quantity of carbon dioxide actually released from solution as well as the bicarbonate buffer capacity of the system.



The carbon dioxide incorporated in the bicarbonate ion is removed from the reactor in the liquid phase rather than the gas phase. There are two factors related to this washout of carbon dioxide. The concentration of alkalinity will control the washout rate at a given retention time. The retention time, or the liquid through-put rate, will also influence the washout rate. Therefore, for a given substrate, digestion at shorter retention times will produce a gas having a higher methane content.

The bicarbonate ion is not the only ion contributing to the alkalinity or buffer capacity. The organic acids and acid salts also contribute to the alkalinity. Equation 10, originally developed by Pohland and Bloodgood (1963) and modified by McCarty (1964b), can be used to calculate the bicarbonate alkalinity.

$$\text{BA} = \text{TA} - (0.85)(0.833) \text{ TVA} \quad (10)$$

where: BA = bicarbonate alkalinity, mg/l as CaCO_3
 TA = total alkalinity, mg/l as CaCO_3
 TVA = total volatile acid concentration, mg/l as acetic acid

The mg/l of acetic acid is converted to the equivalent alkalinity as CaCO_3 by the 0.833 multiplier. The 0.85 factor accounts for the fact that only 85 percent of the volatile acid alkalinity is measured by titration of total alkalinity to pH 4. This equation also assumes that there is no significant concentration of other materials such as phosphates, silicates, or other acid salts which also produce alkalinity.

The above analysis permits an accurate evaluation of the actual bicarbonate ion concentration in a digestion system. Since the normal operating pH in these systems is 7.0 or less, the carbon

dioxide-bicarbonate equilibrium can be combined with the solubility of carbon dioxide to predict the gas composition under various operating conditions. One must also include the effect of the vapor pressure of water. This increases substantially at the thermophilic temperatures. These data are shown in Table 5. The added water vapor in the gas reduces the partial pressure of carbon dioxide which in turn changes the bicarbonate alkalinity equilibrium.

Table 5. Vapor Pressure of Water at Various Temperatures

Temperature °C	30	35	40	45	50	55	60
Vapor Pressure--mmHg	31.8	42.2	55.3	71.9	92.5	118	149.4

Source: *Handbook of Chemistry and Physics*, 40th ed., CRC Publishing Company

Temperature also influences this equilibrium. For a given pH, say 7.0, the ratio of bicarbonate ion to carbonic acid (carbon dioxide) increases. At 30°C this ratio is 4.9 while at 60°C it increases to 5.19. These values are shown in Table 6. When the effect of temperature on the solubility of carbon dioxide as described by Henry's Law is considered, the effect is compounded.

Table 6. Effect of Temperature on K_1 of Carbonic Acid

Temperature °C	30	40	50	60
$K_1 \times 10^7$	4.9	5.04	5.13	5.19

Source: *Handbook of Chemistry and Physics*, 40th ed., CRC Publishing Company

Equation 11 shows this solubility relationship. The change in K is given in Table 7. For a given mole fraction in solution, an increase in temperature will increase the partial pressure.

$$K = \frac{P}{K} = \frac{\text{Partial Pressure} - \text{mm Hg}}{\text{mole fraction}} \quad (11)$$

Table 7 lists Henry's constant for carbon dioxide at various temperatures. The temperature effect on this equilibrium results

Table 7. Henry's Constant for Various Temperatures

Temperature °C	30	40	50	60
K x 10 ⁻⁷	0.139	0.173	0.217	0.258

Source: *Handbook of Chemistry and Physics*, 40th ed., CRC Publishing Company

in a significant reduction in alkalinity. A decrease in the solubility of carbon dioxide by a factor of 0.85 in going from 30°C to 60°C in conjunction with an increase in the vapor pressure of water by a factor of 4.7 produces a significant reduction in the carbon dioxide in solution. This effect is not offset by the changes in the carbon dioxide-bicarbonate relationship with increasing temperature.

The effect of bicarbonate alkalinity, pH and temperature can be calculated from the following equation. This equation will yield

$$P_{CO_2} = K \frac{18 \times 10^{-3} [H^+] [HCO_3^-]}{K_1} \quad (12)$$

where: P_{CO_2} = Partial pressure of CO₂ in gas phase - mm Hg

K = Henry's constant for CO₂

$[H^+]$ = Hydrogen ion concentration - mole/l

$[HCO_3^-]$ = Bicarbonate ion concentration - mole/l

K_1 = Dissociation constant for carbonic acid

the partial pressure of carbon dioxide in a gas which also will have a substantial quantity of water vapor. The gas will be dried prior to analysis or combustion. The partial pressure of carbon dioxide will increase as a result of the elimination of the water vapor. Equation 13 will give the partial pressure of carbon dioxide in a dry gas. This value represents the carbon dioxide composition of the gas as determined by a gas analysis.

$$(P_{CO_2})_{dry} = \frac{760}{760 - P_{H_2O}} (P_{CO_2})_{wet} \quad (13)$$

Temperature has a significant effect on the carbon dioxide-bicarbonate system. Increasing the temperature significantly increases the partial pressure of carbon dioxide for a given pH and alkalinity. These data are shown in Figure 5. At the higher temperatures, considerably more carbon dioxide will be lost from the system in the gas phase. For a given pH, a lower alkalinity will occur at the higher temperatures. More of the carbon dioxide will be lost in the gas rather than be discharged from the system as bicarbonate alkalinity. With more carbon dioxide being discharged in the gas phase, the caustic requirements for pH control would be significantly reduced.

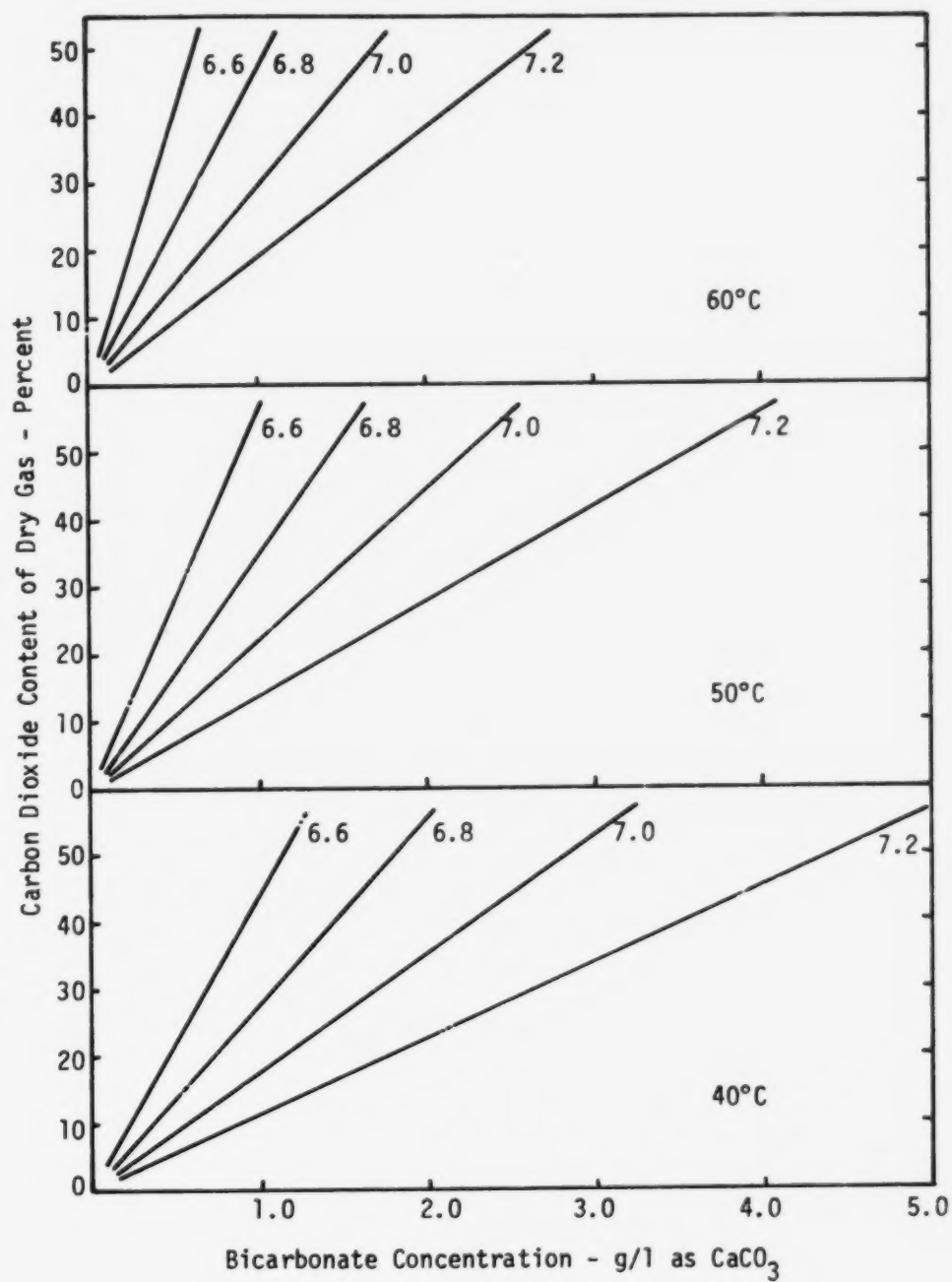


Figure 5. The Relationship Between Gas Composition, Temperature, pH and Alkalinity

EXPERIMENTAL PROCEDURE

Experimental Reactors

Eight reactors of plexiglass construction were built. Each reactor measured 10.5 in. in diameter and 18 in. in height, and had a volume of approximately 24 liters (Figure 6). Except for the reactor top, which was bolted down to the reactor wall and sealed with pliable Permatex, all plexiglass was fused together with ethylene dichloride. Each reactor was fitted with a stainless steel helical worm stirrer equipped at the bottom with a scraper extending almost the whole reactor diameter. A 1.25 in. Bunting oil impregnated bronze bearing fixed to the reactor bottom was used for stirrer alignment. A Garlock Mechanipak seal maintained an airtight joint between the extending 1.25 in. stirrer shaft and the plexiglass top. It also transmitted the weight of the stirrer shaft and the incremental force required for proper operation of the Mechanipak seal to the plexiglass top of the reactor. A Boston Gear flanged cartridge, connected to the stirrer shaft above the Mechanipak seal served for proper alignment of the stirrer shaft and also for transmitting the incremental pressure over and above that supplied by the stirrer weight for proper Mechanipak seal operation.

A 1.25 in. rubber stoppered opening at the top of each reactor served as an inlet for feeding purposes. A 0.25 in. diameter plexiglass nozzle extending 2 in. above the reactor top served to link each reactor to a Tygon tubing gas line leading to a Sargent-Welch precision wet test gas meter. A 1 in. diameter, 1.5 in. long plexiglass nozzle at the bottom of each reactor leading to 1.25 in. diameter Tygon tubing served for effluent withdrawal. A double pipe clamp fixture was used to control the effluent flow through the Tygon tube. Figure 7 is a pictorial view of one battery of the reactors.

Units 1 through 4 were built and operated before units 5 through 8 were completed. The outlet nozzles on the former were located on the cylindrical plexiglass wall about 1 in. above the plexiglass bottom of the reactor. Because of plugging experiences with this type of effluent arrangement, units 5 through 8 featured effluent nozzles extending downward from bottom of the reactor. Reactors 5 through 8 had baffles located in the side walls to improve the mixing by eliminating the circular motion induced by the mixing mechanism.

Each four reactors, 1 through 4, and 5 through 8, formed a battery and had their own framework of Uni-Strut construction and electric motor. The stirrer shafts extending through the reactor tops in each battery were connected by miter gears to an overhead shaft. This latter, in turn, was connected to an electric motor. The

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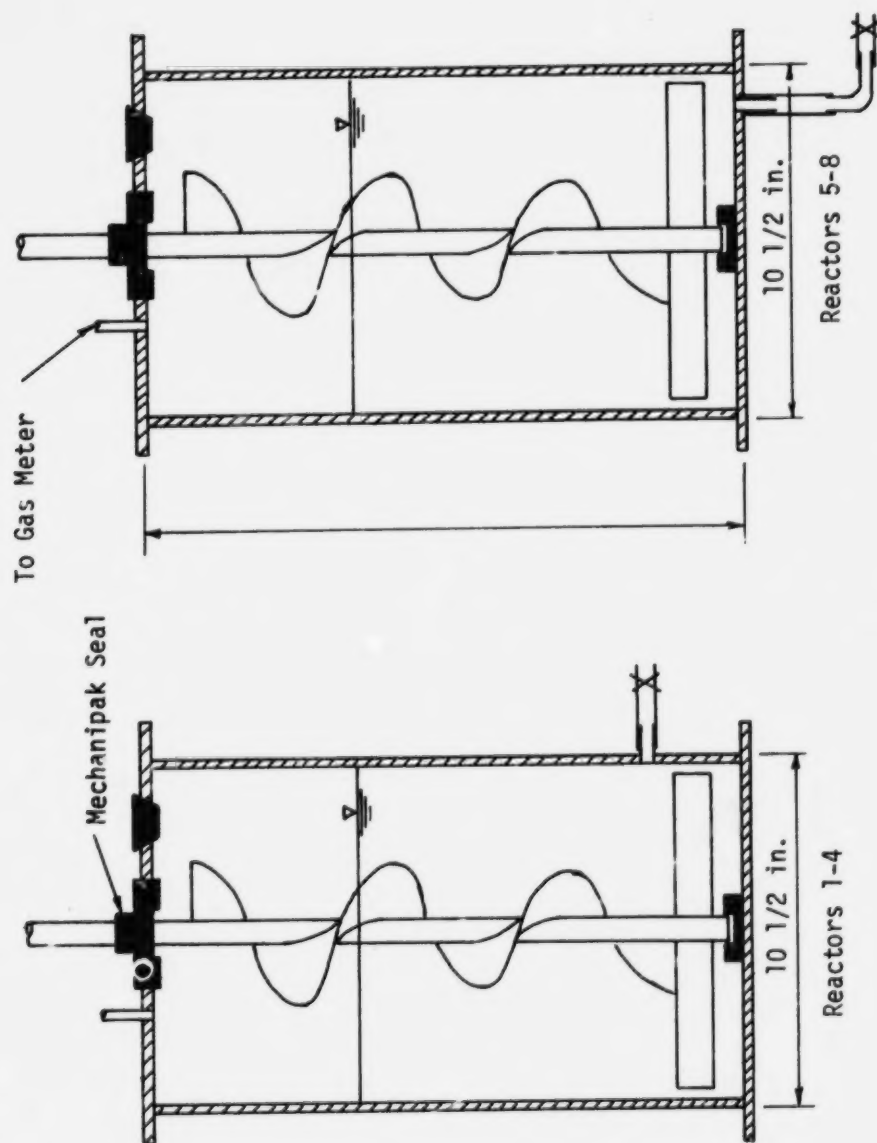


Figure 6. Schematic of Experimental Units

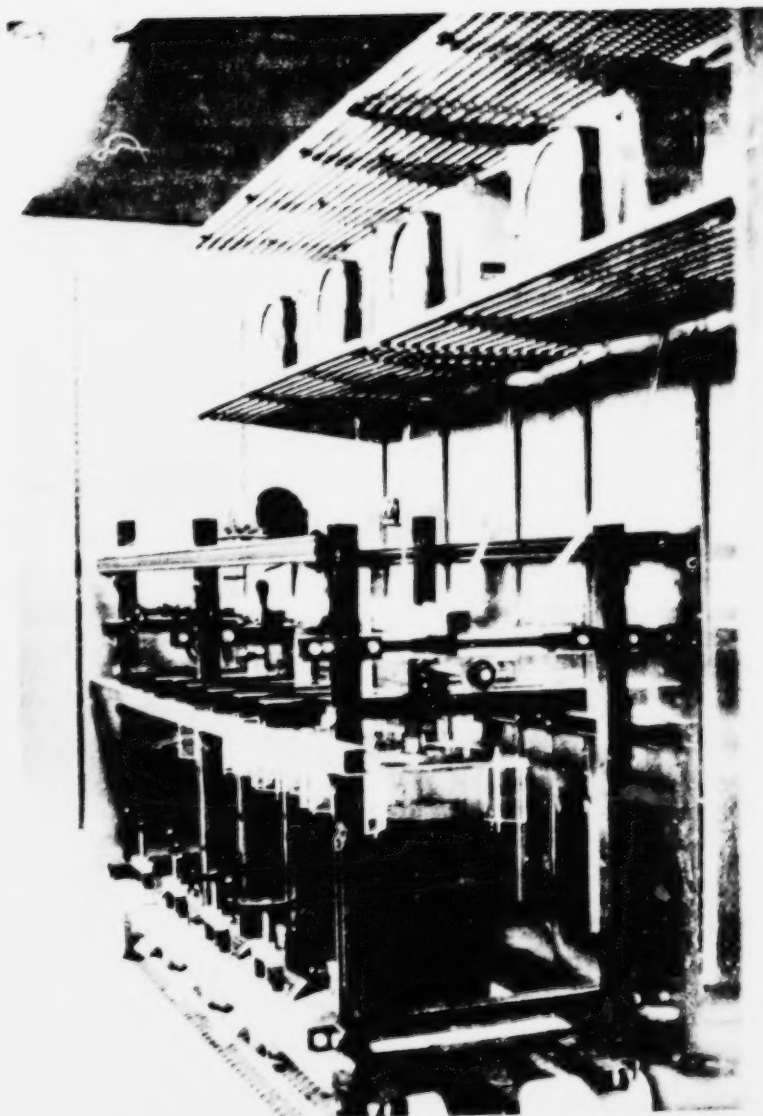


Figure 7. Laboratory Reaction Units (Battery Containing Units 1 - 4)

electric motors were 0.25 horsepower, 60 rpm, Dayton Gearmotors. Both batteries were housed in a constant temperature room in which the temperature could be varied from 35°C to 60°C. All reactors were tested under pressure for gas tightness before operation and the operating volume constituting the reacting contents in each reactor was 15 liters. The stirrers were operated in a continuous fashion at 60 rpm.

Substrate

In the fall of 1969, a trip was made to the Center Hill Research Facility, Environmental Protection Agency, in Cincinnati, Ohio, for the purpose of obtaining a large, homogeneous amount of "typical" domestic refuse that would last the duration of the experimental runs. There, the contents of a collection vehicle serving a residential area were shredded. By inspection, it was determined that the size of the shredded refuse particles was about one inch. It was also noticed that the refuse contained large percentages of dried leaves and branches and paper but contained very little garbage. Also, appreciable amounts of foam rubber, plastic, nylon, styrofoam, etc. were present. No attempt was made to segregate the broken glass and pieces of metal from the refuse before shredding. These latter were also found in appreciable amounts.

Eight 55 gal. drums of refuse were filled, sealed, and all but one placed in cold storage at -20°C. The drum not placed in cold storage served the purpose of providing refuse for waste characterization and later on, for feed to the reactors. Drums were taken out of cold storage one by one as the need for additional amounts of refuse arose. The contents of each drum were allowed to thaw and much of the broken glass and metals was removed by hydraulic separation. Early experiences indicated that the presence of large quantities of broken glass and metal caused excessive wear on the moving parts and frequently caused the mixing mechanism to jam. For a large scale application, it would be possible to design a reactor to resist the wear from these abrasive materials. However, it would be desirable to remove them prior to the digestion step. These materials are inert to biological decomposition and as such there is no point in introducing them into the reactor.

The refuse obtained on the initial trip was inadequate to complete the study. It was necessary to return to the Center Hill Laboratory for an additional supply of refuse in the spring of 1972. This refuse was found to have the same characteristics as the original sample.

Hydraulic separation of the broken glass and metals from the refuse was done in small batches sufficient for about one week's operation of the reactors. A very simple technique was used for the hydraulic

separation. A 30 gal. plastic container was half filled with refuse and the balance of the container was filled with tap water. The refuse and water were manually mixed and allowed to stand for a short period. Styrofoam, foam rubber, plastics and other floating, biologically inert material were removed by hand. The glass, metals and grit settled to the bottom of the container. The degritted refuse was manually placed on a screen which allowed the bulk of the water to drain from the refuse. Although this process was not 100 percent efficient in removing the inert materials, it did reduce the operational problems experienced prior to this degritting step.

The degritted refuse was stored in a covered container in a constant temperature room maintained at 2°C. Frequent moisture determinations were made on this degritted refuse and sufficient amounts of tap water added to obtain the desired solids content in the feed slurry. This feed slurry was added daily on a batch basis. Prior to addition of the slurry, an equivalent volume of reactor contents was withdrawn. This effluent was used for analytical and other determinations. Although the feeding procedure was conducted with care, some air was entrained in the reactor vessel. Gas analysis showed that the oxygen was quickly consumed and, because of the large volume of the reactor contents, no oxygen toxicity was observed.

Nutrients were added directly to the digesters whenever the need for nutrient addition arose. Nitrogen was fed in the form of NH_4Cl and the concentration of ammonia nitrogen was maintained in the 200 to 500 mg/l range. Whenever nitrogen was introduced to a digester, phosphorus in the form of K_2HPO_4 was also added such that a nitrogen to phosphorus ratio of five to one was maintained.

The pH of the reactors was controlled by the addition of NaOH. The pH was maintained in the range of 6.7 to 7.0. When ever the pH dropped to near 6.6, enough NaOH was added to increase the pH to a more desirable level. By experience it was determiend how much alkali was needed to maintain the pH in the desired range. NaOH was added directly to the daily feed slurry, resulting in a relatively constant pH value.

Analytical Procedures

pH

pH measurements of the contents of the reactors were made on a daily basis throughout the duration of this study. These readings were made as soon after withdrawal of the samples as possible using a Beckman Electromate pH Meter.

Alkalinity

Digester effluent was centrifuged in a Sorvall Superspeed RC-2 at 6000 rpm ($G = 5,860$) for 15 to 20 minutes. The supernatant was then used for the alkalinity determination which was carried out according to the procedure described in *Standard Methods* (1965). Instead of using methyl orange to indicate the titration end point as described in *Standard Methods*, the titration was assumed to be complete at a pH of 4.5. All alkalinity reported are in mg/l as CaCO_3 .

Volatile Acids

A two-stage centrifugation procedure was employed. Samples were first centrifuged as described for alkalinity determinations. The supernatant was further centrifuged in the same centrifuge, using a different size head at 15,000 rpm for 15 minutes ($G = 27,000$). The resulting supernatant was then analyzed for volatile acids (total) using the Column Partition Chromatography method as described in *Standard Methods* (1965). All values are reported in mg/l as acetic acid.

Cellulose and Non-Cellulose Carbohydrates

The feed and effluent were regularly analyzed for cellulose and non-cellulose carbohydrates. The Anthrone Method, as described by Viles and Silverman (1949), Gaudy (1962), Snell and Snell (1961), and Koehler (1952), was used for these determinations. This proved to be a very sensitive test requiring much patience and practice. For a step-by-step description of the procedure followed, refer to Appendix A.

Gas Analysis

The gases generated were analyzed for methane and carbon dioxide using a Model 25 V Fisher Gas Partitioner. This instrument is a gas chromatograph designed for the quantitative determination of substances which are in the gaseous state at room temperature. The gas partitioner was coupled to a Model SR Sargent Recorder. The heights of the peaks are used analytically to determine the concentrations of the various gases in the mixture. This is done through comparisons with peak heights obtained with a standard gas mixture.

In this study, a standard gas mixture containing 40 percent carbon dioxide and 60 percent methane by volume was used. The resulting percentages of carbon dioxide and methane in the digester gas were then normalized to 100 percent. A fresh standard was prepared for each day on which gas analyses were conducted. Gas

analyses were not conducted as frequently as desired due to operational problems with the partitioner.

Fixed and Volatile Solids

Fixed and volatile solids determinations were made on a weekly basis on both the feed and digester effluent. These determinations were carried out according to the procedure described in *Standard Methods* (1965) for sludge and bottom sediments with one modification. No water bath was used prior to drying in an oven.

To assure as representative a sample as possible, samples were blended in a Waring Commercial Blender and then poured back and forth in a vigorous manner from one container to another prior to placement in the evaporating dishes. To assure that the loss of volatile solids in the determination of fixed solids is held to a minimum, samples were dried at 70°C and at 103°C, as recommended by *Standard Methods* (1965), and the results compared. It was concluded that no significant loss of volatile matter were encountered with the higher temperature which was subsequently used for all fixed solids determinations.

Ammonia Nitrogen

Weekly determinations of ammonia nitrogen were made on the digesting contents of all units for the purpose of assuring that sufficient concentrations of nutrients were being maintained for optimal digestion. Ammonia nitrogen concentrations were maintained through this study at a minimum level of 100 mg/l and a maximum level of 500 mg/l as nitrogen. In this manner adequate nutrient levels were maintained without raising the ammonia nitrogen level to toxic proportions.

Samples were first centrifuged as described for alkalinity determinations, and the supernatant used for carrying out the analysis for ammonia nitrogen according to the Distillation Procedure described in *Standard Methods* (1965).

Other Analytical Methods Used for Refuse Characterization

These analyses were conducted according to the analytical techniques recommended by the American Public Works Association (*Municipal Refuse Disposal*, 1966) with the following exceptions. The BTU value of the refuse was determined with a Parr-Bomb Calorimeter using the technique recommended by the manufacturer. The carbon, hydrogen and nitrogen analysis was conducted on an F & M Scientific CHN Analyzer. The carbon and hydrogen values were within the sensitivity range of this instrument. However, the nitrogen values were too low to be detected. The wet Kjeldahl

analysis was used to determine the nitrogen content of the refuse. The chemical oxygen demand (COD) of the refuse was conducted according to *Standard Methods* (1965).

Dewatering of Digested Solids

A major step in the processing of organic refuse for energy recovery and for volume reduction is dewatering of the digested solids. The filterability of the digested solids was examined using a filter test leaf technique. This technique has been described by Eckenfelder (1966). The procedure followed in this study is described in Appendix B. As the test leaf and filter cloths used were Eimco products, this company's literature was the one mostly utilized and its instructions followed.

To obtain the various sludge concentrations investigated for dewaterability in this study, the collected digester effluent slurry was allowed to thicken by gravity to the desired concentration.

Although laboratory tests showed that the cake was completely formed in five seconds, 30 seconds was selected for form time for all test leaf runs, this time being considered the minimum practical time for plant scale installations. No washing of the product cake was practiced and the drying time was varied from one to 2.5 minutes in increments of half a minute.

The laboratory vacuum outlets served as a vacuum source and provided a static vacuum pressure of 18 in. of mercury throughout the filter test leaf runs.

Attempts to keep the sludge well agitated during cake formation were made using special magnetic stirring rods. However, these attempts all failed and were consequently abandoned. The procedure followed consisted of thoroughly agitating the sludge mixture with a stirring rod immediately prior to immersing the test leaf. Due to the fact that cake formation was very rapid, in the order of five seconds, this method of agitation is believed to be satisfactory.

Suspended solids determinations of the filtrate were carried out according to the procedure described in *Standard Methods* (1965).

The Eimco test leaf used was circular in design and had a diameter equal to four inches. Filter cloths were fixed to the test leaf with an adjustable clamp band. The first two or three tests with a new cloth were discarded as they are not an accurate indication of fabric performance.

Three filter test cloths, all 1/1 plain weave, and made of polyethylene, were utilized in these dewaterability determinations. All

were Eimco products. Table 8 summarizes the pertinent information for each of these three test cloths.

Table 8. Test Cloth Specifications

Test Cloth Number	Eimco Style Number	Material	Weave	Thread Count
1	PO-801 HF	Polyethelene	1/1 Plain	112 x 48
2	PO-801 RF	Polyethelene	1/1 Plain	105 x 40
3	PO-808	Polyethelene	1/1 Plain	40 x 23

RESULTS

Characterization of the Raw Refuse

Upon acquisition of the refuse, analyses were run for the purposes of defining the characteristics and chemical composition of the raw refuse. These analyses followed hand separation of the metals and glass, and are summarized in Table 9.

Table 9. Composition of Refuse (Initial Determinations)

Determination	Number of Determinations	Average	Range
Volatile total solids	3	81.8%	81.6-82.1
Lipids	3	6.2%	5.98-6.36
Chemical Oxygen Demand	5	0.982 gm/gm	0.943-1.021
Calorific value	2	7510 BTU/lb dry solids	7410-7610
Carbon	2	44.1%	43.68-44.50
Hydrogen	2	5.7%	5.625-5.750
Nitrogen (CHN analyzer)	2	0%	---
Nitrogen (wet Kjeldahl)	4	0.68%	0.666-0.700
Phosphorus	5	0%	---
Total carbohydrates	5	52.75%	51.5-53.5
Cellulose	5	35.8%	35.0-37.0

NOTE: The composition of non-cellulose carbohydrates is the difference between total carbohydrates and cellulose.

During the course of the study, analyses were routinely made on the raw feed refuse. The analyses included total volatile solids, cellulose, and non-cellulose carbohydrate determinations. These analyses were made on the refuse as fed to the reactors and are summarized in Table 10.

It can be seen by a comparison of Tables 9 and 10 that differences exist between the two results. The metals and glass in the initial analyses had been removed by careful hand sorting as opposed to the less efficient hydraulic separation technique employed for preparation of the feed for the reactors. This accounts for the refuse solids data shown in Table 9 being

Table 10. Experimental Determinations on the Feed Refuse

Determination	Number of Determinations	Average Result	Standard Deviation
Total volatile solids	25	73 %	± 1.70 %
Cellulose	44	22.3%	± 0.63 %
Non-cellulose carbohydrates	46	1.0%	± 0.113 %

81.8 percent volatile as opposed to only 73 percent for the data in Table 10.

The cellulose determinations in the earlier series of experiments were made on samples from which most of the fixed solids and non-combustibles had been removed. The samples used for the latter determinations were identical in composition to reactor feed and contained an average 73 percent volatile solids. When this difference in the volatile solids content of the sample is taken into account, the results for cellulose presented in Tables 9 and 10 are comparable, within acceptable experimental and refuse characteristic variation. However, the non-cellulose carbohydrate determinations cannot be reconciled even if differences in the total volatile solids content of the samples are taken into consideration. If the samples used in the initial analyses had contained 73 percent volatile solids instead of 81.8 percent, the adjusted non-cellulose carbohydrate content would be substantially greater than the one percent non-cellulose carbohydrate content of the reactor refuse feed.

The non-cellulose carbohydrates measured by the procedure described earlier consist mainly of starches, sugars, etc. These organic compounds are easily biodegradable and are metabolized by a host of aerobic, anaerobic and facultative microorganisms. It is possible that subsequent to the removal of the refuse drums from cold storage, sufficient biological activity occurred to metabolize these compounds. Because of the relatively long service life of each drum, sufficient time was available for breakdown of the starches and sugars even at the low moisture content of the refuse in the drums. Growth of fungi was noticed on the refuse. Short-chain organic acids, alcohols, aldehydes, etc. are known to be intermediaries in the anaerobic decomposition of glucose. Under anaerobic conditions, the starches and sugars would have been transformed into a form not measurable by the non-cellulose carbohydrate procedure and may partially explain the discrepancy in the two results. Aerobic stabilization may also have occurred, generating carbon dioxide and water from these compounds.

The non-cellulose carbohydrates measured by the anthrone procedure are water soluble. As such, a large portion of these compounds would be expected to go into solution during the hydraulic separation procedure used for preparation of the feed for the reactors. Since the majority of the water used for hydraulic separation was discarded, most of the solubilized non-cellulose carbohydrates were undoubtedly discarded. This would also contribute to the discrepancies in the non-cellulose carbohydrate data as seen in Tables 9 and 10.

Refuse constituents not analyzed for include, among others, hemicelluloses and lignins. Hemicellulose, a cellulose-like carbohydrate is only partially measured by the cellulose procedure described earlier. The pentose portion of the hemicelluloses, together with lignin and its derivatives, various organic resins, pitch, proteins, plastics, rubber, nylon, etc., undoubtedly constituted the balance of the refuse volatile constituents not shown in Table 9.

Phase I

The first phase of this study was designed to determine if refuse could be effectively converted to methane gas by the anaerobic fermentation process. Upon completion of the construction of the reactors, the experimental work was initiated. Two problems became apparent soon after the reactors were started. The first was mechanical in nature. The shredded refuse contained significant quantities of glass and metal. This material caused excessive mechanical problems and it became necessary to remove it prior to digestion. A simple hydraulic separator was employed to remove over 90 percent of the glass and metal. Most of the dust and other fine inert particles were not removed by this separation step. This material is not biodegradable and simply occupies space in the reactor. High mixing velocities and associated high power requirements would be necessary to prevent the deposition of this material in the reactors. Consequently, an early decision was made to separate essentially all of the glass and metal from the shredded refuse before introducing it to the reactors.

The second problem was one of acclimation of the microorganisms to the refuse. Actively digesting sludge was obtained from the local water pollution control plant to use as a seed for the reactors. The initial operation of the process was poor. Rapid stabilization did not occur as was expected. Careful control of the pH and organic acids prevented failure of the system. After a four week acclimation period, the process began to respond; the organic acid level began to decrease and gas production was substantially increased. The reactors continued to efficiently digest the refuse for the balance of this phase of the project.

Refuse is known to be nutrient deficient, having an excessively high carbon to nitrogen and carbon to phosphorus ratio. Active biodegradation requires supplemental nutrients. Inorganic salts of nitrogen and phosphorus were added to the reactor feed. Effective conversion of the refuse to methane was obtained in this system.

Every municipality has quantities of wastewater sludge that must be processed. This sludge contains nitrogen and phosphorus as well as micronutrients. The effect of the addition of small quantities of raw sewage sludge on the process was evaluated. The effect is shown in Figure 8. All four reactors were receiving 50 grams of refuse (dry solids) per day. Two of the reactors were also receiving three grams (dry solids) of raw sewage sludge.

The gas production from the reactors with sewage sludge addition was disproportionately higher than the other two. At a 15-day retention time, the addition of sewage sludge increased the gas production 24 percent. The addition of sewage sludge at the 30-day retention time only increased the gas production by 12 percent. However, the gas production at both retention times was significantly greater than that which would be accounted for by the increased loading of six percent resulting from the addition of three grams of sewage sludge. All subsequent studies included the addition of sewage sludge to all reactors.

Immediately after the initiation of the feeding schedule, problems were encountered with maintaining the contents of the various reactors at a pH greater than 6.6. Initially this was not unexpected in view of the high concentrations of short-chain organic acids that very quickly developed. However, even after the reactors achieved stable operating conditions and the concentration of the short-chain organic acids dropped to 100 to 300 mg/l as acetic acid, it was still necessary to add caustic in order to hold pH above 6.6. This was accomplished initially with the addition of sodium bicarbonate. Because of the release of CO_2 upon addition of NaHCO_3 to the low pH reactor contents, surges in gas production were experienced after addition of NaHCO_3 for pH adjustment. To eliminate this problem, NaOH was used for the balance of the study. Enough caustic was added to raise the pH of the particular digester to 6.7 to 6.8. During this phase of the study no further additions of alkalinity were made until the pH of the digester under consideration dropped to near 6.6. Due to their higher rate of washout, the caustic requirements were observed to be higher at the lower retention times. The caustic requirements at various retention times were determined in Phase 2.

Once the systems stabilized, the alkalinity values for the various reactors remained relatively stable for the duration of the six week

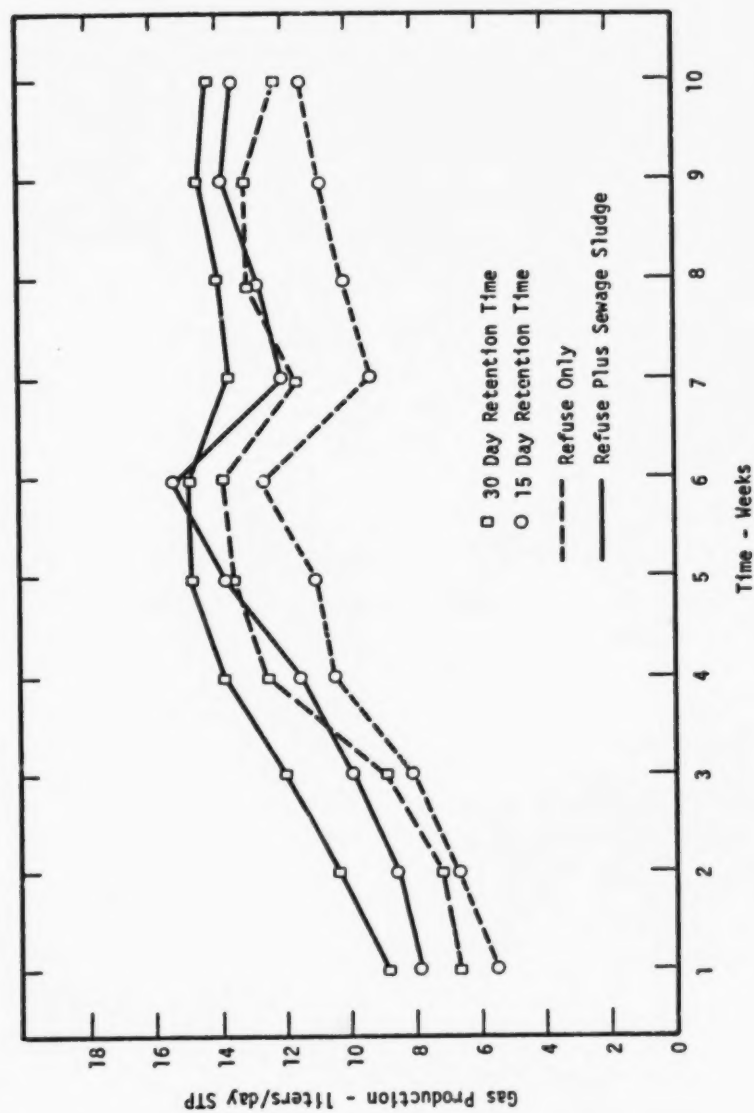


Figure 8. Effect of Raw Sewage Sludge Addition on Refuse Digestion

experimental run. As can be seen in Table 11, alkalinity increased with an increase in the concentration of feed solids but, even then, the observed values are considerably lower than typical alkalinity values observed in properly operating municipal sludge digesters. It should be noted that the alkalinities of the reactors receiving 100 ml of raw primary sewage sludge in addition to refuse were higher than those receiving only refuse.

Table 11. Summary of Operating Conditions at 35°C - Phase 1

Reactor No.	Retention Time-days	Volatile Solids Feed - gms/day	Gas Production - STP		Alkalinity mg/l as CaCO ₃
			l/day	scf/lb V.S. Add.	
1	30	8.76	3.13	5.71	1275
2	15	8.76	2.79	5.10	1200
3	30	18.25	6.31	5.54	1500
4	15	18.25	5.44	4.77	1400
5	30	36.50	12.89	5.65	1700
6	15	36.50	10.88	4.77	1750
7*	30	38.70	14.40	5.95	1900
8*	15	38.70	13.55	5.61	2000

* Reactors receiving three grams of raw sewage sludge solids (100 ml of sludge) per day.

The need for the addition of caustic to prevent a precipitous drop in pH, coupled with the low alkalinities present in the reactors, clearly demonstrates that the characteristics of the refuse were such that adequate buffer did not develop from its decomposition. The nitrogen content of the refuse was very low indicating that very little protein was present. The deamination of protein is a major source of alkalinity in anaerobic fermentation.

A summary of the results of Phase 1 is shown in Table 11. These data were collected over a six week period during which the system was at equilibrium. The gas production from the digestion of refuse only averaged 4.84 and 5.63 SCF at a 15-day and 30-day retention time, respectively. As would be expected, the longer retention times resulted in a higher gas yield. The addition of sewage sludge did increase the gas production indicating some benefit associated with this material. The cause for this increase is probably related to two factors. Sewage sludge contains a host

of micronutrients in a form readily available to the microorganisms. Also, the continued introduction of the wide variety of microorganisms present in sewage sludge may promote the development of a more optimum culture for stabilization of the organic material. Whatever the reason, there is an advantage in adding sewage sludge when digesting refuse. Since this sludge is readily available in most, if not all, communities, it would be a logical addition in the digestion of refuse.

The composition of the gas produced during Phase 1 was measured and these data are shown in Table 12. These values are for the

Table 12. Composition of Gas Produced - Phase 1

Reactor Number								
Methane %	61	62.5	58.5	60	58	58	59.5	60
Carbon Dioxide %	39	37.5	41.5	40	42	42	40.5	40

dry gas. There is very little variation in the quality of the gas produced by these reactors. Since the composition of the gas is controlled by the pH-alkalinity relationship, these data reflect the variation in these two parameters. The addition of caustic for pH control can be expected to change the quality of the gas produced by this process.

Solids Destruction - Phase 1

The solids concentrations fed the various reactors and the measured effluent concentrations from these reactors along with the percent volatility of the influent and effluent solids are shown in Table 13. All values are six-week averages during which time the units were in equilibrium. During this period, the effluent solids exhibited relatively constant quantitative and qualitative characteristics. Also shown in Table 13 are the apparent reductions in volatile and total solids.

Based on the volatile solids destruction efficiencies and the six-week average daily gas productions, the volume of gas produced by the various reactors is shown in Table 14. The values are expressed at standard temperature and pressure.

The values observed in Table 14 range from 8.41 to 10.92 cu ft/lb volatile solids destroyed. With the exception of the 10.92 cu ft/lb

Table 13. Apparent Volatile and Total Solids Reduction

Parameter	Reactor							
	#1	#2	#3	#4	#5	#6	#7	#8
Feed solids conc.	2.4	1.2	5.0	2.5	10.0	5.0	10.6	5.3
%								
Volatility of feed solids, %	73	73	73	73	73	73	73	73
Effluent solids conc., %	1.00	0.73	2.19	1.32	3.34	2.18	3.00	2.16
Volatility of effl. solids, %	62.3	64.1	61.8	67.5	72.7	74.6	75.6	74.6
Apparent volatile solids red., %	64	47	63	51	67	55	71	58
Apparent total solids red., %	58	39	56	47	67	56	72	59

Table 14. Apparent Reactor Gas Production Per Unit Volatile Solids Destroyed

Gas Production Units	Digester							
	#1	#2	#3	#4	#5	#6	#7	#8
l/gm volatile solids destroyed	0.555	0.683	0.550	0.583	0.529	0.538	0.526	0.601
cu ft/lb volatile solids destroyed	8.88	10.92	8.80	9.30	8.46	8.60	8.41	9.60

volatiles destroyed that reactor 2 exhibited, all values are well below the theoretical gas production of 11 cu ft/lb volatile solids destroyed. This wide variation in gas production per mass of volatile solids destroyed is not justifiable. The refuse feed solids introduced to all digesters were of identical chemical composition and anaerobic fermentation, as such, should yield nearly

identical gas productions per unit mass destroyed. A major reason for these variations is associated with the problem of obtaining a representative sample of the reactor effluents. A true solids balance could not be obtained on the reactors. This is evidenced by the fact that not all the fixed solids fed the reactors could be accounted for in the effluents from these reactors. This point is illustrated by the data in Table 15.

Table 15. Fixed Solids Balance

Digester No.	Influent Fixed Solids - g/day	Effluent Fixed Solids - g/day	Fixed Solids Lost - g/day
1	3.24	1.89	1.35
2	3.24	2.51	0.73
3	6.75	4.18	2.57
4	6.75	4.29	2.46
5	13.50	4.57	8.93
6	13.50	5.50	8.00
7	14.30	3.70	10.60
8	14.30	5.50	8.80

The loss of fixed solids in the digestion process is not an indication of biological breakdown since these solids are not significantly changed in this process. The problem is attributed to two considerations. First, the nature of the refuse made sampling very difficult. Second, visual observation showed a significant amount of material entwined with the mixing mechanism.

Since one cannot account for all of the fixed solids, it is also logical that one cannot account for all of the volatile solids either. The apparent destruction of volatile solids which are not actually converted to gas results in lower gas production values on a per unit mass of volatile solids destroyed basis. This explains the low values of gas production shown in Table 14. Reactor 2 appeared to have the best solids balance with only an average of 0.73 gms of fixed solids not accounted for on a daily basis. The calculated gas production of reactor 2, at 10.92 standard cu ft/lb volatile solids destroyed is the highest among all reactors, and closely approximates the predicted production of 11 standard cu ft/lb solids destroyed. It, therefore, may be concluded that a gas production of 11 standard cu ft/lb of volatile solids destroyed is a realistic value for the refuse used in this study.

Since the gas production should be constant and equal to about 11 cu ft/lb of volatile solids destroyed for all reactors, and because of the difficulties encountered in obtaining a true solids balance, the destruction of volatile solids can be calculated on the basis of gas production. These calculated reductions are shown in Table 16. They are more uniform and are typical of what may be expected from sewage sludges. Reflected in these values also are the increased volatile solids destructions in the reactors receiving sewage sludge in addition to refuse. Such increases in the destruction of volatile solids are accompanied by increased gas yield.

Table 16. Adjusted Volatile Solids Destruction

Digester No.	Volatile Solids Destruction, %
1	52
2	46
3	50
4	43
5	51
6	43
7	54
8	51

The difficulty encountered in obtaining a true solids balance precludes the expression of gas production in terms of volatile solids destroyed. All remaining gas production data will be expressed in terms of volatile solids added.

Phase 2 - Mesophilic Temperature Studies

The results from the first phase of the study indicated that conversion of the organic refuse to methane was indeed possible. No inhibition of the digestion process was experienced with the refuse used in this study. It was necessary to add major nutrients as the refuse was essentially devoid of nitrogen and phosphorus. It was observed that the addition of sewage sludge to the feed stock significantly improved the biological breakdown of the refuse. It was also observed that the natural buffering capacity of the process was inadequate to maintain the pH at a satisfactory level. It is necessary to add caustic to the reactors to maintain the pH in excess of 6.6.

With these observations as a background, a second series of experiments were undertaken to establish the rates of stabilization of the refuse at 35°C. The feed stock for the reactors was the same as previously employed with the exception that all reactors received 50 ml of raw sewage sludge. The organic load to all units was held constant at 50 grams of dry refuse plus 50 ml of sewage sludge (1.5 grams dry solids). This produced a loading of 0.21 lb solids per cubic foot of reactor capacity. The loadings of the units are given in Table 17.

Table 17. Reactor Loadings - Phase 2

Reactor No.	Retention Time Days	Feed Solids Concentration %	
		Total	Volatile
1	4	1.37	0.94
2	4	1.37	0.94
3	6	2.10	1.44
4	8	2.74	1.87
5	10	3.43	2.35
6	15	5.15	3.53
7	20	6.86	4.70
8	30	10.30	7.06

As with the previous series, feeding was done daily on a batch basis. There was initial concern about the shock effect of withdrawing and adding one-fourth of the volume of reactors 1 and 2. However, they were not adversely affected by this operation. The shift to these lower retention periods was made gradually using the active digesting units from the previous series of experiments. The transition to these retention times was smooth and the organic acids remained below 1000 mg/l in all units.

After equilibrium was established, the average operating conditions for a five-week period were as given in Table 18. Frequent alkalinity and volatile acid determinations were made during this run. The results are shown in Appendix C, Table 35. Because of the relative minor variation in these analyses, they were conducted much less frequently in the remaining runs. As long as reactor operation was normal, these analyses were conducted simply to determine the level of the parameter in each reactor.

The pH was controlled by the addition of caustic which also contributed to the alkalinity. The caustic requirements are expressed as meq per 15 liters per day. There is an excellent

Table 18. Reactor Operating Conditions - 35°C

Reactor Number	pH	Alkalinity mg/l CaCO ₃	Volatile Acids mg/l Acetic	Caustic Added meq/day
1	6.77	1400	420	93
2	6.78	1460	420	93
3	6.78	1700	380	59
4	6.75	1840	324	56
5	6.75	1900	278	56
6	6.77	2300	282	38
7	6.81	2600	279	38
8	6.87	3000	330	38

correlation between alkalinity and retention time. The long retention times are associated with a high concentration of solids, both suspended and dissolved. The natural buffer capacity at the longer retention times is much greater as evidenced by the decrease in the amount of caustic required to maintain the pH at an acceptable operating level. This consumption of caustic may mitigate against the use of short retention time in the practical application of the process. It should be indicated that the caustic consumption will be a function of the chemical characteristics of the refuse being processed. Ammonia is one of the primary ions in developing a natural buffer in the anaerobic process. Therefore, any refuse, such as a typical municipal refuse, that is very low in nitrogen may be expected to exert a significant demand for additional buffer.

The use of actual refuse as a substrate was a cause of substantial variation in the data collected. Figure 9 shows the variation in daily gas production for 33 consecutive days when the reactors were considered to be operating under equilibrium conditions. These values are the meter readings and they have not been adjusted to standard temperature and pressure. Reactors 4 and 8, having longer retention times, had less variation in the daily volume of gas generated. Some of the variations can be attributed to variation in feed. When all three curves show similar trends, this trend would result from the feed. Only 50 grams of dry solids (shredded domestic refuse) were added to each reactor. In selecting the refuse feed for daily feeding from the storage container, it was possible to introduce variations. In a similar manner, preparation of a batch of refuse by hydraulic separation could result in variations in feed characteristics. This would result in increases or decreases in gas production over a several day period.

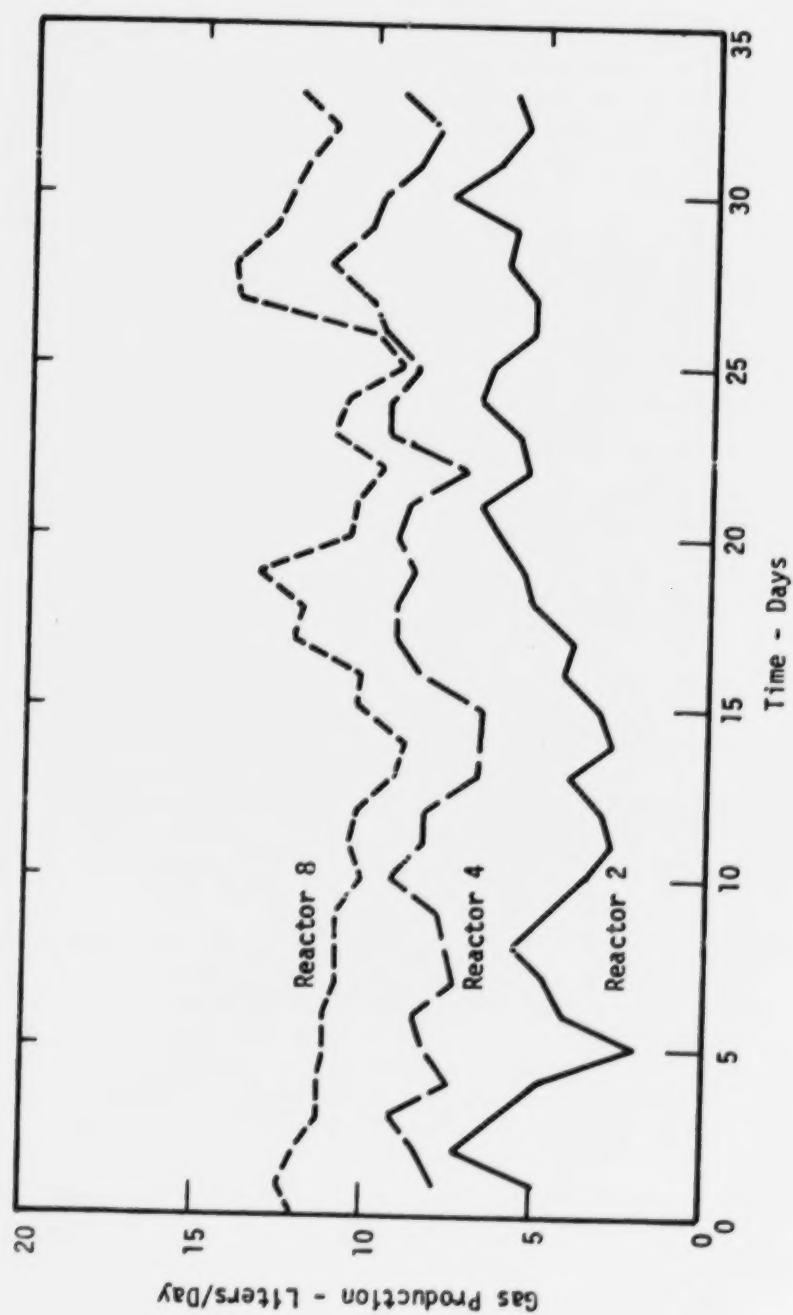


Figure 9. Variation in Gas Production Under "Equilibrium" Conditions - 35°C

Variations between reactors can also result from the selection and weighing of the individual 50 grams for each reactor.

Additional variation may be expected from reactor instability, in particular the units operating at the short retention times. Other problems such as incorrect reading of the gas meters can add to the variation. The gas meters recorded cumulative gas productions, so a high reading followed by a low reading, or the reverse, may be due to meter reading error.

Upon completion of the 35°C test conditions, the temperature of the system was gradually increased to 40°C. The equilibrium operating conditions for a five-week period at 40°C are given in Table 19.

Table 19. Reactor Operating Conditions - 40°C

Reactor Number	Retention Time Days	pH	Alkalinity mg/ℓ CaCO ₃	Volatile Acids mg/ℓ Acetic	Caustic Added meq/day
1	4	6.77	1244	278	80.3
2	4	6.77	1319	240	80.3
3	6	6.78	1595	222	70.3
4	8	6.78	1780	298	52.0
5	10	6.78	1982	405	49.0
6	15	6.80	2356	642	36.0
7	20	6.83	2554	733	23.0
8	30	6.87	3087	866	19.0

The pH and caustic additions were recorded daily. However, the frequency at which the alkalinity and volatile acid analyses were conducted was reduced considerably from the 35°C study. These individual results are given in Appendix C, Table 36. There was no significant variation in these parameters during the steady state condition. Therefore, only a few analyses were conducted for the purpose of fixing the level of these parameters under these specific conditions.

The pH was maintained at the same level as for 35°C. However, the caustic required to maintain this pH level was significantly less than that required at the lower temperature. The alkalinity was slightly lower in the 40°C study, especially at the shorter retention times. The longer retention times show essentially

the same level of alkalinity at both temperatures. However, the volatile acids in reactors 5 through 8 were significantly higher than reactors 1 through 4. These organic acid salts do add to the alkalinity. Therefore, correcting the alkalinity for that contributed by the organic acids would give the actual alkalinity due to the bicarbonate ion, which would be substantially less than at the 35°C temperature.

The reduction in caustic requirements is a result of the temperature effect on the carbon dioxide-bicarbonate relationship as previously discussed. More carbon dioxide is being removed from the system in the gas phase rather than in the liquid phase as bicarbonate ion.

The higher level of organic acids measured in reactors 5 through 8 can only be related to the differences in mixing between these two sets of reactors. The baffles on the sides of these reactors appeared to have an adverse effect on mixing when the concentration of the solids in the reactors was high. Visual observations showed areas in the reactor that were very poorly mixed. This poor mixing is apparently responsible for the higher level of organic acids found in reactors 5 through 8.

This problem was not experienced in all cases. The volatile acid data in Table 18 do not show any significant variation between the two sets of reactors. Therefore, this mixing problem is not exclusively related to the reactor characteristics. It was complicated by the buildup of material on the mixing mechanism. The plastics, cloth and other fibrous materials would entwine around the mixer. It was necessary to remove this material periodically. However, the reactors having a concentrated feed slurry experienced a rapid buildup of this material. Consequently, the problem was most pronounced in reactors 5 through 8.

This observation does suggest that mixing conditions must be given careful consideration in a large scale system. The degree of mixing required is based upon maintenance of homogeneity within the reactors. The rate limiting step appears to be the hydrolysis of the fibers. Therefore, mass transfer considerations do not enter into the mixing considerations.

The variation in gas production for 34 consecutive days is shown in Figure 10. Reactor 2 shows variations similar to that found at 35°C. However, the reactors with the longer retention times and higher solids content showed somewhat greater variations, particularly near the end of the period. Some of these variations may be due to the feed variations. Mixing in reactor 4 and 8 may be a cause for the more extreme variations. These variations are not unreasonable when one considers the characteristics of the substrate employed.

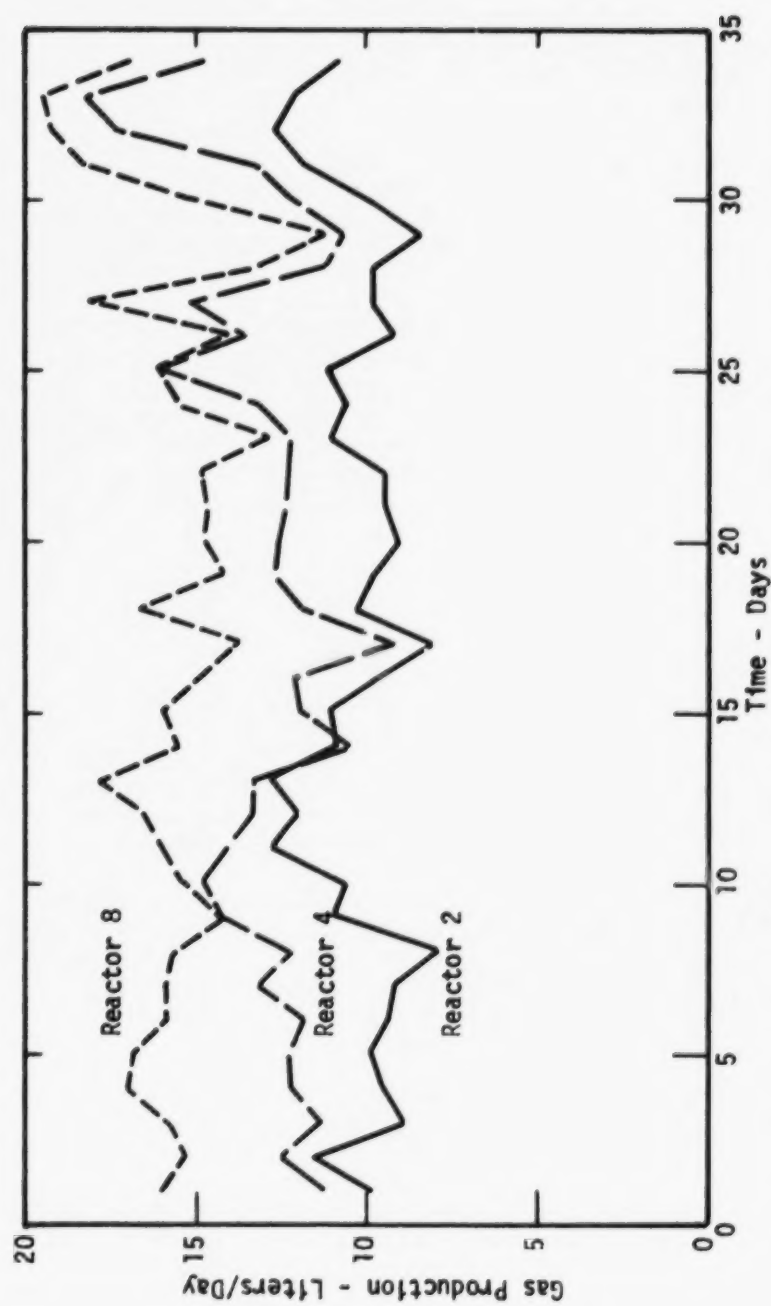


Figure 10. Variation in Gas Production Under "Equilibrium" Conditions - 40°C

When the temperature was increased to 45°C, the alkalinity and caustic requirements decreased slightly. These data are shown in Table 20. The average for the alkalinity and volatile acids represent only four analysis. Because of the inhibition expected and observed at this temperature, only limited data were collected. The volatile acids remained at a low value at this temperature even though the gas production decreased substantially. The gas production will be discussed in detail later. The acids produced from the breakdown of the organic material were efficiently fermented. It appears that the reduced gas production resulted from a decrease in the breakdown of the complex organic materials.

Table 20. Reactor Operating Conditions - 45°C

Reactor Number	Retention Time Days	pH	Alkalinity mg/l CaCO ₃	Volatile Acids mg/l Acetic	Caustic Added meq/day
1	4	6.73	1269	143	77.3
2	4	6.67	1214	531	78.6
3	6	6.74	1312	116	57.0
4	8	6.78	1612	81	50.7
5	10	6.78	1752	171	42.1
6	15	6.81	2002	282	35.7
7	20	6.80	2335	220	27.0
8	30	6.82	2410	221	20.0

The variation in gas production is shown in Figure 11. As with the previous temperature studies, considerable day-to-day variation was experienced. Similar trends were exhibited by all three reactors shown, suggesting that much of the variation is due to the variation in feed characteristics. There are some variations between individual reactors which can be attributed to the factors previously mentioned.

Gas Production and Composition - Phase 2

The gas production and composition is shown in Table 21. There is a very rapid increase in the quantity of gas produced for a given loading with increasing retention time until about 15 days. At this retention time gas production continues to increase, but at a very slow rate. While a 30-day retention time will yield more gas, it may be of questionable value to operate at these

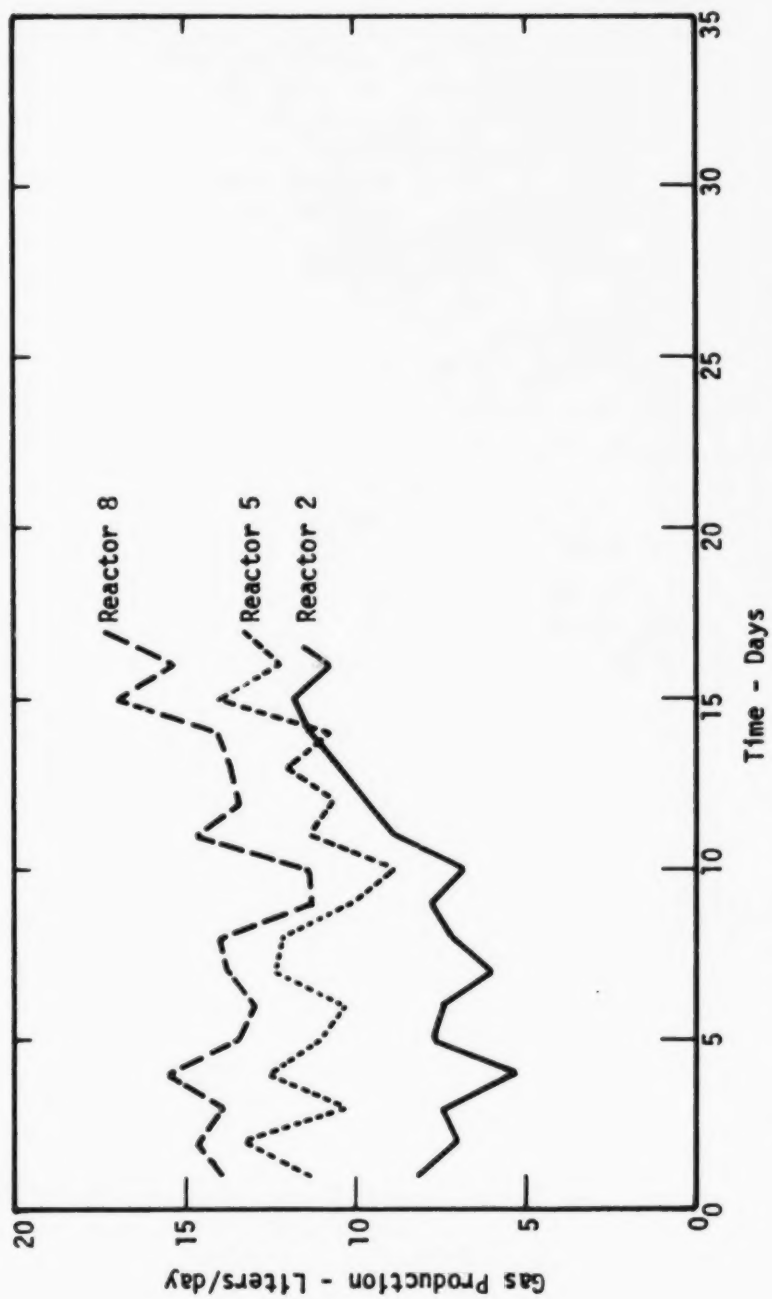


Figure 11. Variation in Gas Production Under "Equilibrium" Conditions - 45°C

Table 21. Gas Composition and Production - 35°C

Reactor Number	Retention Time Days	Methane %	Carbon Dioxide %	scf Gas lbs V.S. Add.
1	4	69.7	30.3	2.02
2	4	69.8	30.2	2.02
3	6	64.3	35.7	2.91
4	8	58.6	41.4	3.36
5	10	57.2	42.8	3.84
6	15	53.8	46.2	4.23
7	20	53.4	46.6	4.28
8	30	53.8	46.2	4.39

longer periods due to the substantial increase in capital costs for larger reactors.

The composition of the gas was not particularly encouraging at the longer retention times as shown in Table 21. In the units receiving large amounts of caustic, a substantial amount of the carbon dioxide produced reacted with the caustic to produce bicarbonate alkalinity.

Since gas production was low in these units, enough carbon dioxide could be absorbed in the liquid to significantly lower the carbon dioxide content of the gas. At the high gas production rates, this effect would not be significant because of the larger quantity of gas produced as well as the lesser amount of caustic added.

The effect of temperature on gas production is shown in Figure 12. A significant increase in gas production results from increasing the temperature from 35°C to 40°C. This effect is observed at all retention times. A continued increase in temperature to 45°C causes a reduction in gas production. The optimum temperature for the mesophilic micro-organisms appears to be near 40°C. The temperature was gradually increased when the 40°C study was completed. A temperature increase of 0.5°C to 1.0°C every week was used to determine the maximum mesophilic temperature. Gas production increased until a temperature of 43°C was exceeded. The gas production significantly decreased as the temperature was increased. Therefore, it was concluded that 40°C to 42°C was the optimum temperature for mesophilic fermentation of organic refuse.

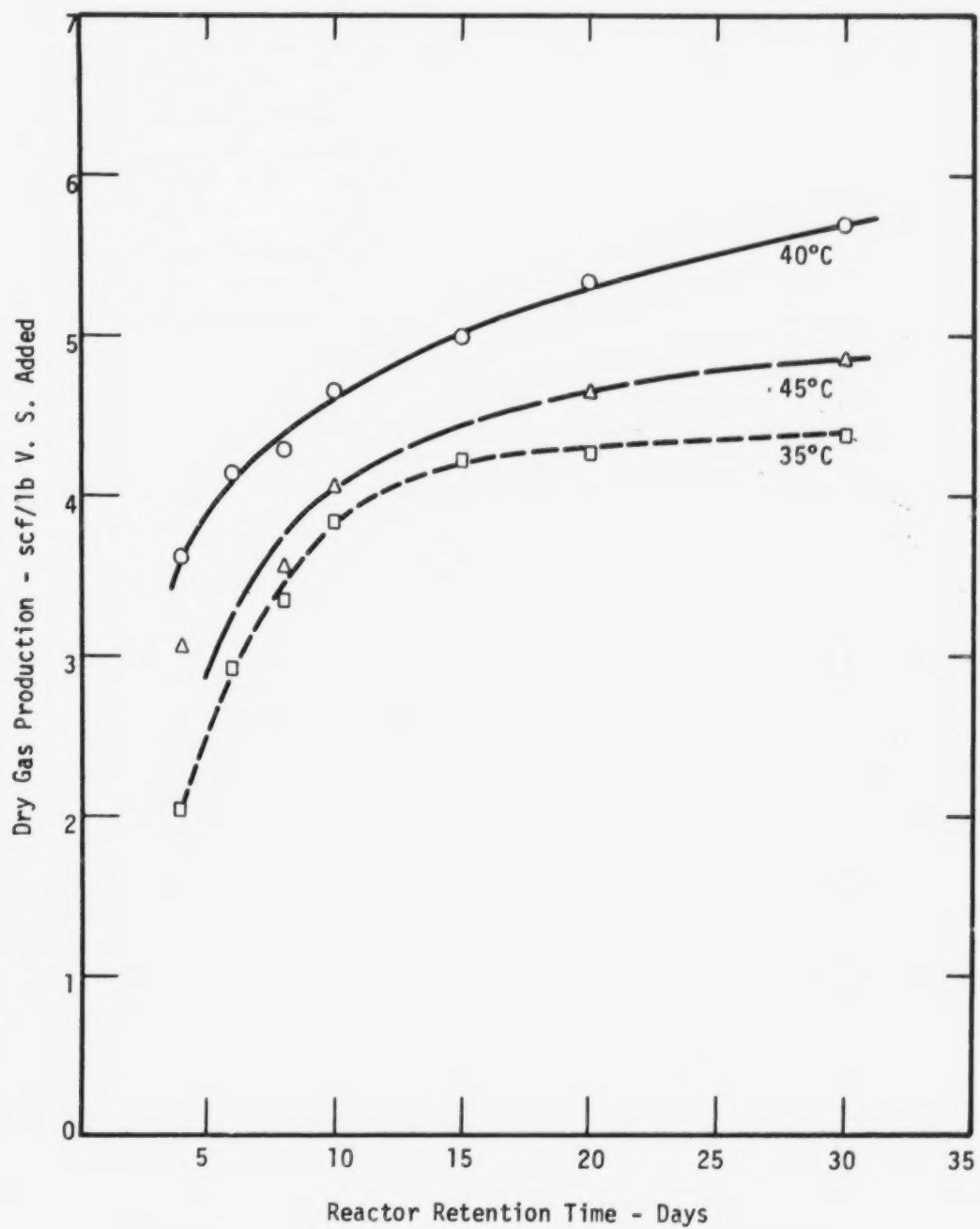


Figure 12. Gas Production at Mesophilic Temperatures

Phase 3 - Thermophilic Temperature Studies

This phase extended the temperature into the thermophilic range. After completion of the mesophilic temperature studies, the temperature was increased to 50°C. All temperature changes were made gradually in order to avoid temperature shock. The reactors had been operating for about two months at 45°C. Even with the gradual increase in temperature, there was a lag in the initiation of thermophilic digestion. After a four week period the gas production was stabilized.

After the acclimation period, data were collected on the operation of these reactors at 50°C. The average conditions during this six week period are shown in Table 22. The individual analyses from which the average alkalinity and volatile acids were calculated are given in Appendix C, Table 37. As with the previous studies, the pH was controlled by the addition of caustic. Increasing the temperature to 50°C reduced the alkalinity and caustic requirements to levels below those found in the mesophilic range (see Tables 18, 19 and 20). The reactors operated effectively at this temperature without any problems.

Table 22. Reactor Operating Conditions - 50°C

Reactor Number	Retention Time Days	pH	Alkalinity mg/l CaCO ₃	Volatile Acids mg/l Acetic	Caustic Added meq/day
1	3	6.75	1156	292	92
2	4	6.78	1154	210	87
3	6	6.83	1251	187	54.7
4	8	6.81	1394	257	52.5
5	10	6.82	1544	218	45.0
6	15	6.85	1778	325	22.5
7	20	6.86	2082	302	15.0
8	30	6.89	2343	402	12.5

The gas production was substantially increased, but was extremely variable. Figure 13 shows 35 consecutive days of gas production for these reactors. During the initial period of equilibrium, the gas production increased substantially with retention time. All reactors exhibited variations that can be attributed to variation in feed characteristics. Reactor 2 was considerably more erratic during the first 15 days. Toward the end of the period, the gas production from reactor 2 was essentially the same as

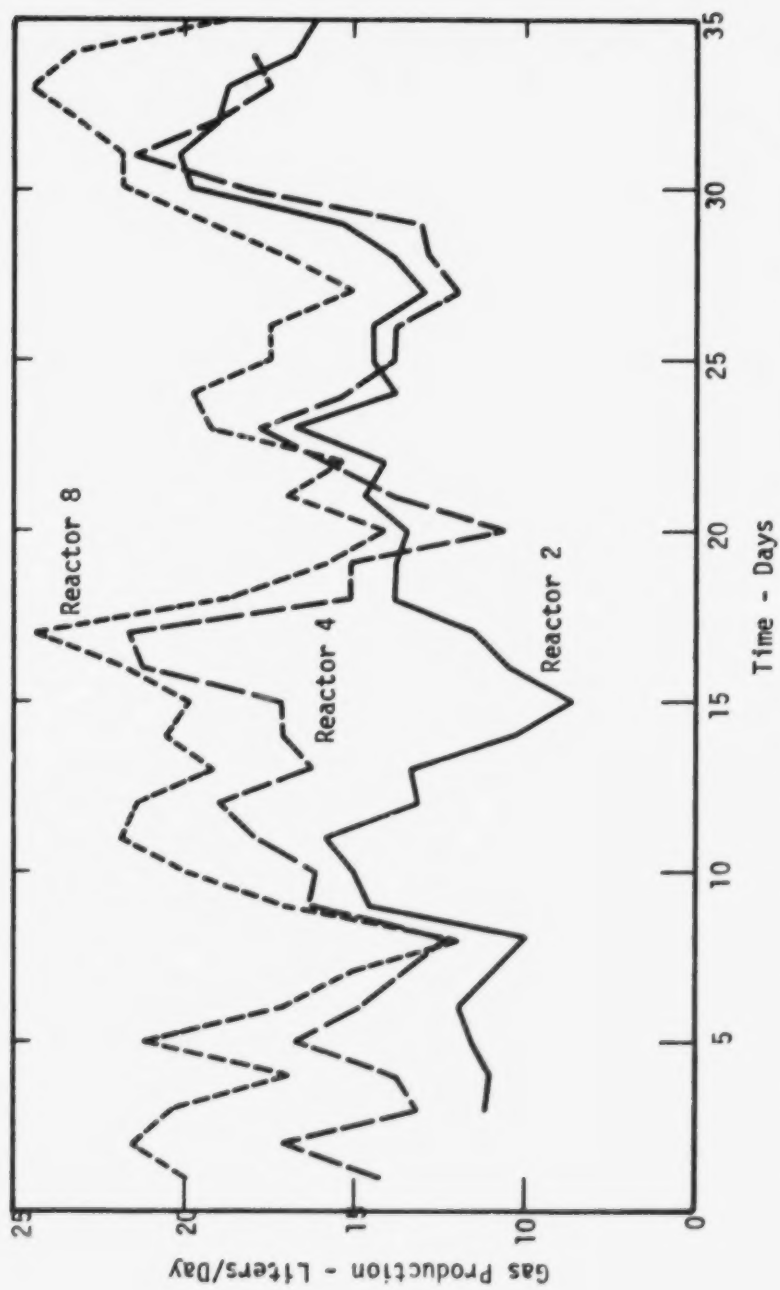


Figure 13. Variation in Gas Production Under "Equilibrium" Conditions - 50°C

the gas production from reactor 4. The difference among all three reactors was considerably reduced. It may be that additional time was required to acclimate the shorter retention time reactors. It does appear that the longer retention times yield little additional gas production at the higher temperature.

Upon completion of the data collection at 50°C, the temperature was gradually increased to 55°C. The average operating conditions for this eight week period are shown in Table 23. The individual analyses used to calculate the average alkalinity and volatile acids are given in Appendix C, Table 38. The average alkalinity

Table 23. Reactor Operating Conditions - 55°C

Reactor Number	Retention Time Days	pH	Alkalinity mg/l CaCO_3	Volatile Acids mg/l Acetic	Caustic Added meq/day
1	3	6.85	1050	112	63.7
2	4	6.85	1300	58	58.5
3	6	6.85	1395	64	37.4
4	8	6.85	1515	79	30.0
5	10	6.85	1435	208	23.7
6	15	6.85	1545	215	10.7
7	20	6.90	1490	298	3.7
8	30	7.00	1885	332	1.1

was reduced by the temperature increase. The caustic required to maintain the pH was substantially reduced, especially at the longer retention times. Reactors 7 and 8 operated at a pH of 6.9 and 7.0, respectively, and required a limited quantity of caustic to maintain this pH. It is probable that these reactors could have operated successfully without the addition of any caustic.

The possible effect of mixing can also be seen from the level of volatile acids found in reactors 5 through 8 (see Table 23). These reactors were not as well mixed as the other four. The organic acid concentration was three to four times greater than that in reactors 1 through 4. These reactors with the shorter retention time could be expected to have higher organic acid levels because of the shorter time available for the biological stabilization to take place. This higher organic acid level would not significantly decrease the gas production, but it would suggest that some factor may be limiting the biological process. Since the

degree of mixing and feed slurry concentration were the only differences between these two sets of units, it must be concluded that these differences were affecting the gas production.

The variation in daily gas production is shown in Figure 14. The variation in feed characteristics appears to be a major cause in the variability of the gas production since all three reactors show essentially the same variation. The gas production from all three reactors was essentially the same. Therefore, it appears that there is little increase in gas production with increasing retention times at the higher temperature.

The thermophilic studies were completed with the 60°C temperature run. This cut off was selected because it is near the upper limit of the thermophilic temperature range. Studies of mixed cultures at temperatures of 65°C (149°F) would expect thermal inactivation of many organisms. The previous studies of anaerobic digestion show a complete inactivation at 65°C (Golueke, 1958). The average operating conditions for this seven week study are shown in Table 24. The data from which the average alkalinity was computed was based on only two analyses. Only one analysis for volatile acids was conducted, indicating that they were within the normal range. Since no operating problems were encountered, additional determinations were not made because of the limited project staff available during this period of the study.

Table 24. Reactor Operating Conditions - 60°C

Reactor Number	Retention Time Days	pH	Alkalinity mg/L CaCO_3	Caustic Added meq/day
1	3	6.8	1030	9.14
2	4	6.8	1140	7.85
3	6	6.8	1530	6.43
4	8	6.9	1600	5.57
5	10	6.9	1920	3.00
6	15	6.8	1900	1.86
7	20	6.8	1900	0.86
8	30	6.8	2030	0.29

The alkalinity again decreased with the temperature increase. The caustic requirements were substantially reduced. It would appear that a very limited amount, if any, caustic would be required for

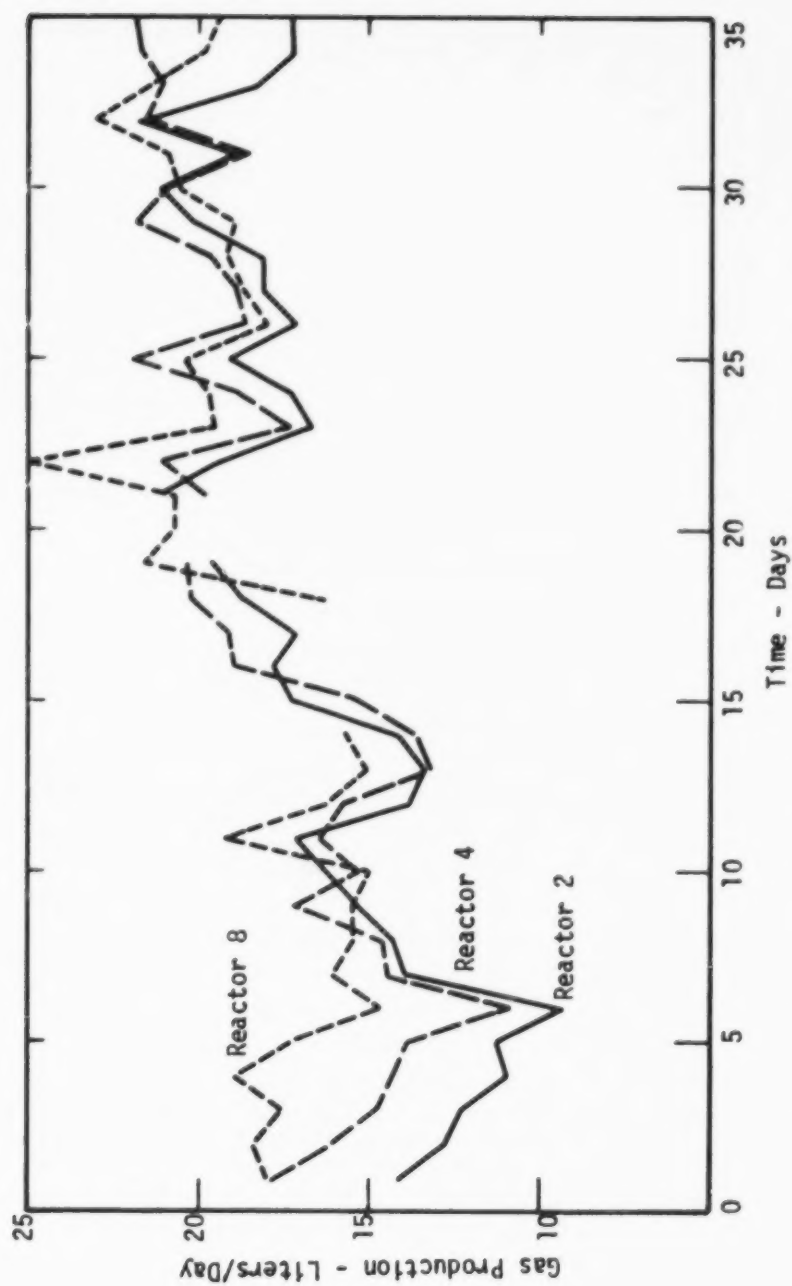


Figure 14. Variation in Gas Production Under "Equilibrium" Conditions - 55°C

pH control at the 60°C temperature. The pH levels were above the minimum recommended value of 6.6 in all reactors.

The daily gas production is given in Figure 15. As with the other runs, considerable variation was experienced in the gas production. During the first 18 days, all these reactors were producing essentially the same quantity of gas. There was considerable variation between the reactors during the balance of the time period. There was no obvious reason for this variation. Part of it may be attributed to variation in the feed.

Gas Production and Composition - Phase 3

The gas production and composition for 60°C is given in Table 25. While the gas production greatly exceeds that of the mesophilic temperature range, the methane content is slightly lower, ranging from 55 to 62 percent as compared to a 53 to 70 percent range at 35°C. With lower bicarbonate concentrations in the liquid phase, more carbon dioxide will appear in the gas phase. The methane content of the gas will decrease with increasing temperatures.

Table 25. Gas Composition and Production - 60°C

Reactor Number	Retention Time Days	Methane %	Carbon Dioxide %	scf Gas	
				lbs	V.S. Add
1	3	62	38	1.96	
2	4	58	42	6.17	
3	6	57	43	6.75	
4	8	57	43	6.16	
5	10	57	43	6.64	
6	15	56	44	6.61	
7	20	55	45	6.90	
8	30	55	45	7.20	

The gas composition is primarily controlled by the alkalinity, temperature and pH. The data in Table 26 show the relationship between pH, alkalinity and gas composition at 60°C. The data represent the conditions on the specific day on which this set of analysis was conducted. The data from reactors 1 and 2, and 5 and 6 show that an increase in pH of 0.15 to 0.2 units will increase the methane in the gas by about 1.5 percent. In a similar manner, data from reactors 3 and 7 show that a substantial increase in alkalinity of 1000 mg/l as CaCO₃ may increase the carbon dioxide

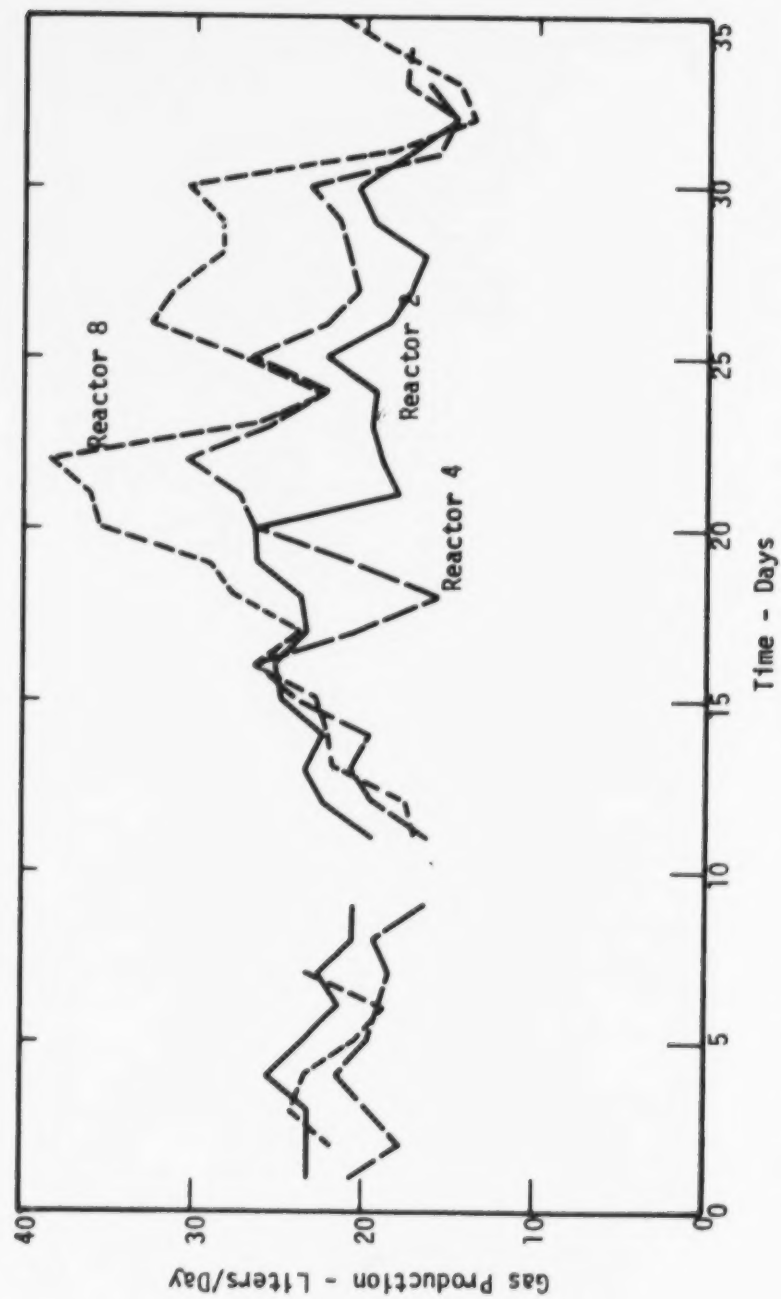


Figure 15. Variation in Gas Production Under "Equilibrium" Conditions - 60°C

Table 26. Effect of pH and Alkalinity on Gas Composition

Reactor Number	pH	Alkalinity mg/l CaCO_3	Methane %	Carbon Dioxide %
1	6.60	1105	53.5	46.5
2	6.80	1125	55.0	45.0
3	6.90	1572	51.9	48.1
4	6.80	1600	50.2	49.8
5	6.85	1920	52.3	47.7
6	7.00	1895	53.5	46.5
7	6.90	2600	49.1	50.9
8	7.05	2030	53.7	46.3

in the gas by only about three percent. The gas composition is much more sensitive to pH shifts than to changes in the alkalinity. This should be expected as a change in pH represents a change in the log of the hydrogen ion concentration.

Figure 16 shows the gas production of 50°C, 55°C and 60°C. The data at the 50°C temperature yielded a reasonably good curve fit. However, the data at 55°C have more variation. The gas data for 60°C provides a somewhat better fitting curve than the 55°C data. However, variations such as evidenced by the 60° data were inherent in this study because of the small reactors and the small daily substrate addition.

Mixing may have been a factor in the variation in the gas production. Reactors 1 through 4, operating with retention times of three to eight days respectively, had significantly better mixing. This was a result of a more dilute feed as well as a different mixing system. The first four data points at 55°C in Figure 16 fit the curve well as shown by the long dashed line. The solid line is an attempt to obtain the curve of best fit for all 55°C data points. These data do show a significant increase in gas production at 55°C, particularly at the shorter retention times. Increasing the temperature to 60°C provides an additional increase in the gas production. Again, the increase is larger at the shorter retention times. Retention times substantially less than 30 days will yield essentially all of the gas that can be generated by the system.

The improved gas production at 55°C and 60°C does not agree with data reported by Golueke (1958). These data are shown in Figure 17. The 55°C and 60°C temperature produced less gas than the 50°C

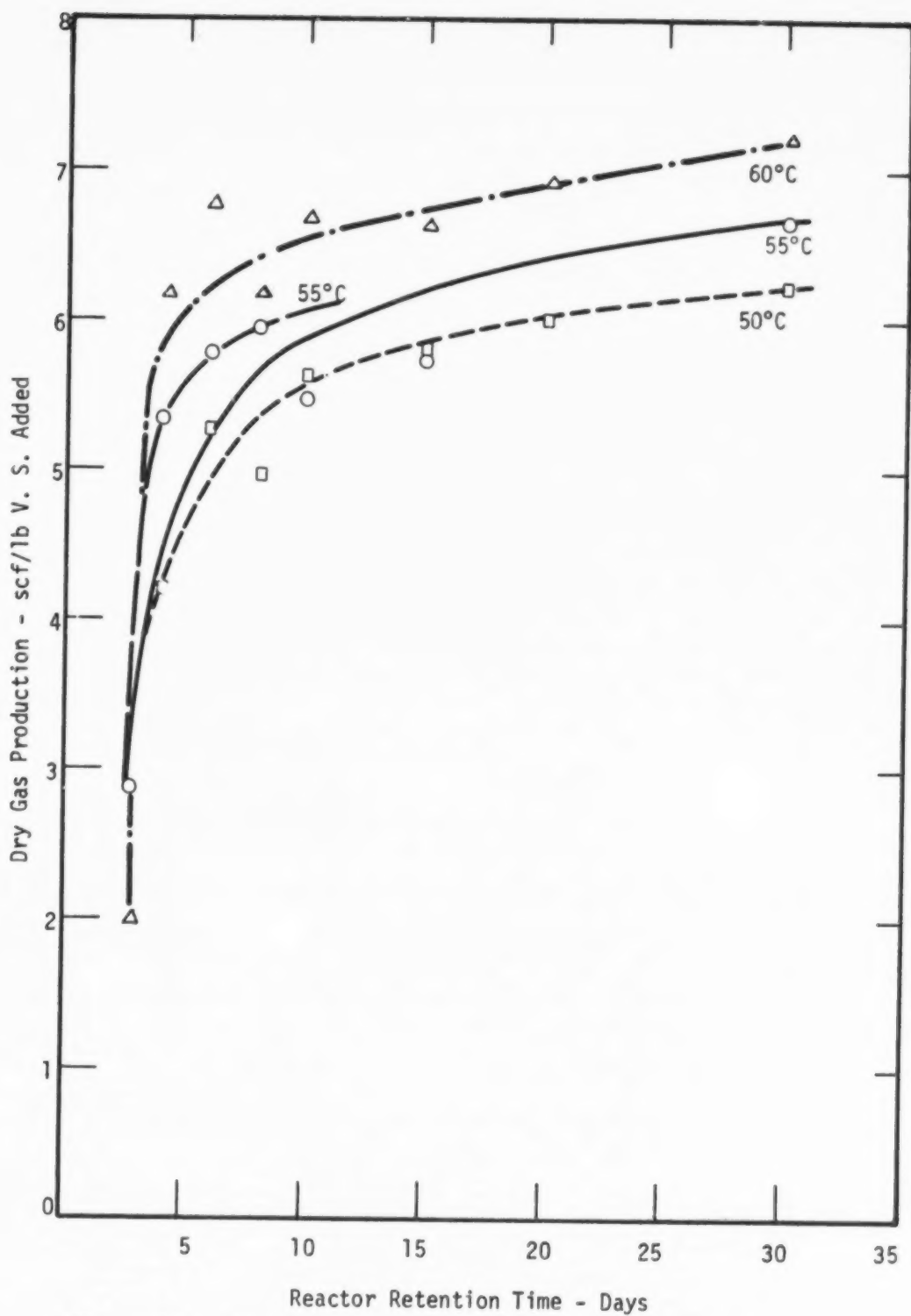


Figure 16. Gas Production at Thermophilic Temperatures

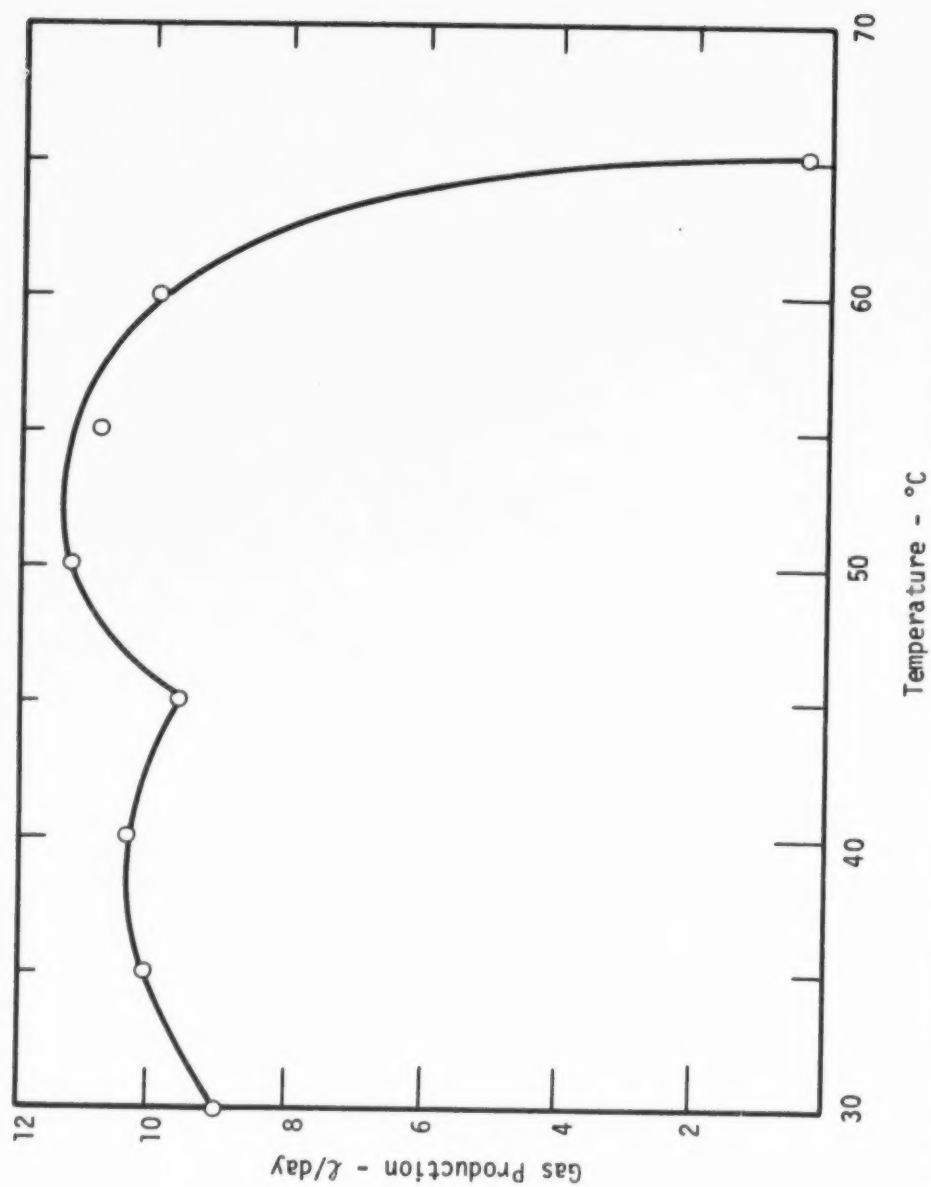


Figure 17. Effect of Temperature on Gas Production (After Golueke, 1958)

temperature. This may have resulted from too rapid an increase in the operating temperatures. In the Illinois study, temperature increases were accomplished at the rate of 1.5°C to 2.0°C per week. This provided adequate time for the microorganisms to acclimate to the increased temperature.

Cellulose Decomposition

The importance of cellulose destruction in the conversion of refuse to methane has previously been discussed. A series of experiments were devised to determine the effect of temperature and pH on this initial rate of cellulose destruction. These studies were conducted in flasks operated anaerobically on a batch basis. The initial slurry concentration was five to six percent. This slurry consisted of raw refuse plus tap water. A starter slurry was added at an amount equal to 25 percent of the total flask contents. This starter solution provides the necessary cell count to insure cellulose hydrolysis. It was obtained from the reactors being fed the domestic refuse. The temperature of the tests were matched with the reactor temperatures so that the starter was acclimated to both the substrate and temperature. Before each temperature run, sufficient quantities of refuse, water and starter were mixed to insure the same initial conditions for each pH level tested at that temperature. This material was stored at 4°C for use in subsequent tests at various pH levels. The Anthrone technique was used to determine the cellulose concentration. The pH was controlled by buffering with sodium bicarbonate and acid, or the addition of sodium hydroxide at the higher pH levels. The pH was constantly monitored with combination electrodes installed in the flasks. Necessary base was added as frequently as required to maintain the pH within ± 1 unit of the desired pH level.

The results of this study are shown in Figure 18. The maximum rate of cellulose utilization was obtained from the slope of the plot of cellulose remaining versus time. The time interval between measurements of the cellulose content in the flasks was 24 hours during the first days of the test and 48 hours thereafter. The effect of temperature may not be as accurate as desired due to variability in the starter solution for different temperature runs. However, the pH effect is accurate since all pH levels at a given temperature were initiated from the same substrate and starter. As can be seen from this figure, the cellulose utilization rate was markedly improved at the elevated pH levels. A pH of 7.5 was the maximum tested. Operating a continuous system in excess of 7.5 would have a significant caustic demand which would markedly increase the operating costs.

Cellulose destruction was monitored in the continuous reactors. The pH of the reactors was maintained at 6.7 ± 0.1 . The data

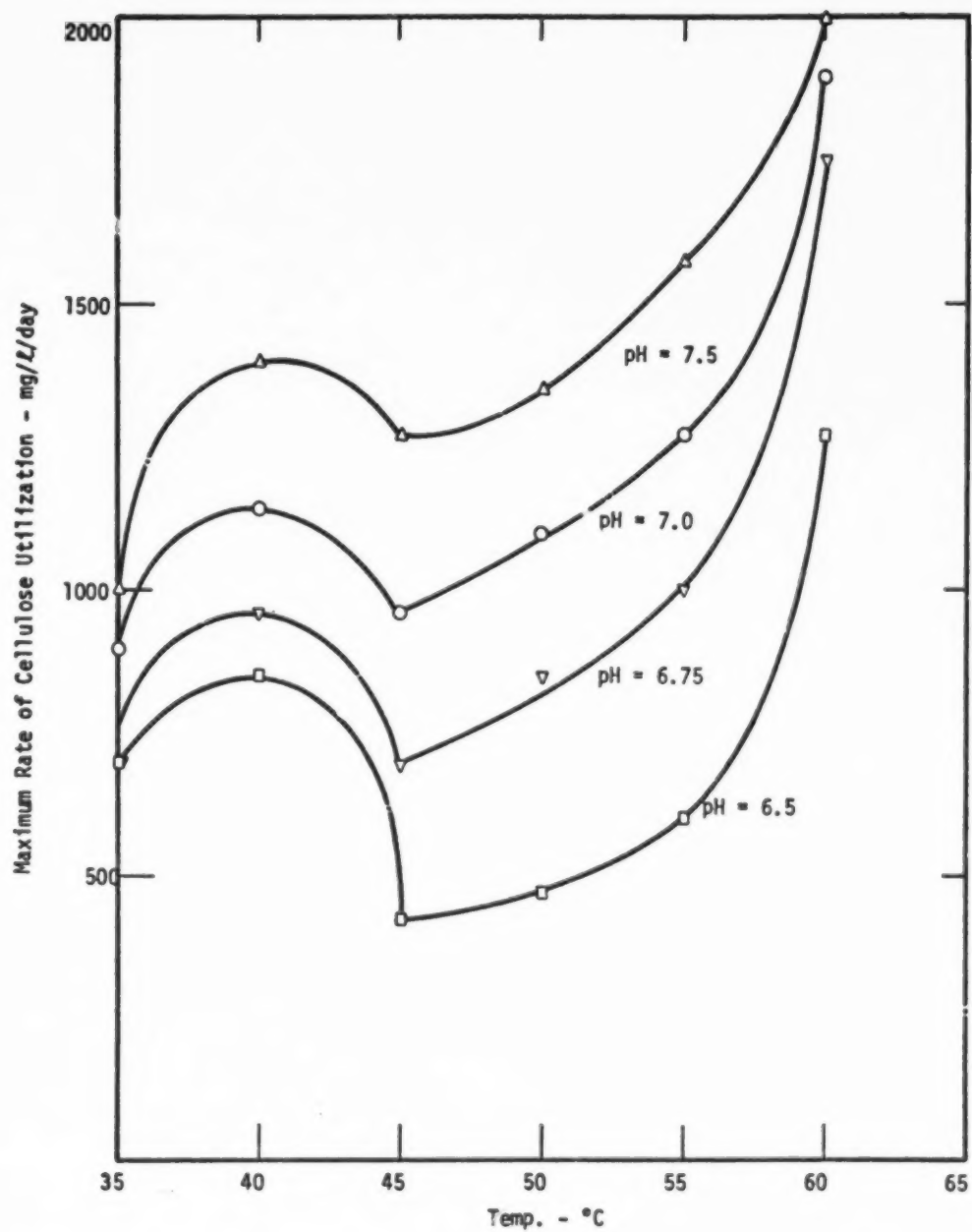


Figure 18. The Effect of Temperature and pH on the Maximum Rate of Cellulose Utilization

shown in Figure 19 were obtained at the 35°C temperature. Even at the lowest retention time, significant hydrolysis of the cellulose occurred. At these short retention times, only traces of noncellulose carbohydrates could be detected. The conversion of the hydrolysis products of cellulose to methane and carbon dioxide was rapid and complete with only small quantities of organic acids remaining.

The 30-day retention time produced a 74.4 percent reduction in cellulose. Again the conversion of the products of cellulose hydrolysis to gas was complete. The rate limiting step in the digestion of refuse appears to be the hydrolysis of the complex organic solids. Data from studies on the digestion of sewage sludge show a significant increase in the organic acid concentration at a retention time of four days as compared to longer retention periods. However, the organic acids were essentially constant in all digestion units. This would suggest that the organic acids were metabolized as rapidly as they were produced at all retention times used in these experiments.

The effect of increased temperatures on cellulose utilization was simply to increase the percent utilization at the lower retention times. At 40°C, the cellulose utilization was 60 percent at the four day retention time, increasing to 70 percent at the 30-day retention time. From these data it would appear that the cellulose remaining in the system is resistant to biodegradation. The available cellulose is utilized rapidly in the thermophilic temperature range.

Dewatering of Reactor Residue

A significant step in the processing of organic refuse for energy recovery will be the dewatering of the digested sludge. This residue must be adequately processed for disposal. The filterability of the digested solids was examined using a filter test leaf technique. A small filter section was immersed in the effluent from the reactors with a vacuum applied for a 30 second form time. The test leaf was withdrawn from the slurry and allowed to dry with the vacuum continuing for 1.0 minute. The filter test cloths were 1/1 plain weave polyethylene with different size pores. A form time of 30 seconds was selected because that would be the minimum practical time. However, the lab tests showed that the cake was completely formed in about five seconds. Since no chemical conditioning was employed in these tests, it may be possible to significantly increase the filter yield by adding chemicals or polymers.

The filter yield at the lower feed slurry concentrations was quite low but was markedly improved at the higher feed slurry

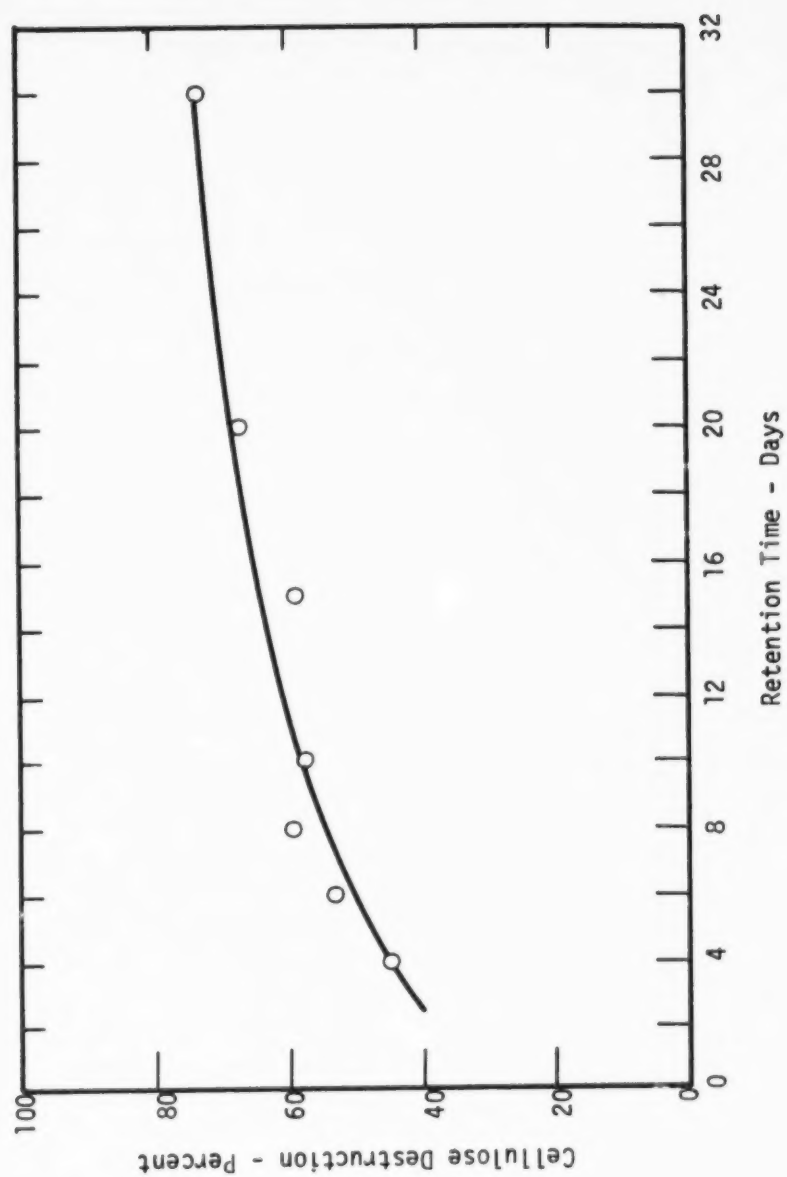


Figure 19. Relationship of Cellulose Destruction with Retention Time at 35°C

concentrations. These data are shown in Table 27. The digested solids appeared to thicken readily. However, with the small volumes of slurry available in the laboratory no meaningful thickening data could be obtained. It does appear that significant improvements in the filter yield can be obtained with thicker slurries. It should be mentioned that the solids concentration in the effluent from the reactor receiving a feed containing ten percent solids varied between 3.0 and 3.5 percent solids. Therefore, one can expect a rather dilute slurry discharged from the reactors.

Table 27. Filterability of Reactor Residue

Test Leaf	Feed Slurry Solids Concentration - %		
	3.5	6.8	9.3
Filter Yield - lbs/ft ² -hr			
1	4.5	13.2	19.0
2	5.0	13.5	19.2
3	5.0	12.5	17.5
Moisture Content - %			
1	75.0	73.0	75.0
2	75.0	77.0	75.5
3	77.0	72.0	76.0

The thickening and dewatering characteristics of this digested material will have to be evaluated on a larger scale. The use of chemicals and/or polymers to improve the dewatering characteristics should also be examined. It does appear from these limited tests that this material has good dewatering potential.

Characteristics of the Separated Inorganic Component

The material removed by the hydraulic separation process consisted primarily of glass, sand and broken ceramics. It also contained some aluminum and ferrous metals. A disposal method for this material must be provided. Particle size was highly variable ranging from particles passing through a 60 mesh sieve to particles an inch or so in size. The hydraulic separation was effective in separating only inert material. The material contained less than ten percent volatile solids.

The bulk density of this material was found to be 2080 lb/cy. The material, as separated from the refuse, was oven dried and no attempt was made to compact the material except for that occurring through normal handling. The weight of a known volume was taken to determine the density. The same material was found to have an average specific gravity of 2.35. As previously mentioned, glass, sand and grit accounted for a major portion of the separated inorganics. A refuse having more metals may generate an inorganic fraction somewhat different than that produced from this refuse.

Digestion of Hydropulped Refuse

The refuse used in this study was shredded to a nominal particle size of one inch. The rate limiting step in conversion of the organic refuse to methane is the hydrolysis of the fibers. This rate is surface area dependent. The surface area in turn can be related to the particle size. In order to determine if hydropulped refuse would improve the conversion rate, a quantity of hydropulped refuse was obtained from the Black Clawson demonstration project at Franklin, Ohio. The separated material was obtained from this facility in a dry form.

Visual examination of the hydropulped material indicated that fibers had been separated initially. However, in the drying process, the fibers tended to form mats. Therefore, the apparent particle size was not significantly smaller than the refuse being used in the study. Upon addition to the reactor, the fibers may have separated readily. An analysis of this material indicated that much of the inorganic material had been removed from the refuse. The material was 87.7 percent volatile.

The hydropulped refuse was used as the substrate for the reactors for about two weeks at the end of the study. The temperature was 60°C. The reactors did not show any noticeable difference in the gas production or operating conditions. The gas production exhibited the usual variation, but the average for the period was the same as when the substrate was the shredded refuse. The long retention times and mixing in the reactor may tend to negate any benefit of the hydropulping. In the reactor, fiber separation is very complete. It would appear that, in fact, the hydropulping operation is accomplished in the reactor. Therefore, increased rates of conversion should not be expected with the dry hydropulped feed. This substrate did not contain the organic material that may have been leached from the refuse during hydropulping. Therefore, a direct feed of the hydropulped slurry to the reactor may yield a different response than the dried feed. This would be a result of the presence of the leached organics rather than a smaller fiber.

PROCESS EVALUATION

Based on the results of this research, an analysis of the process potential can be made. This process evaluation has been divided into two sections, materials balance and production costs for the product gas. The material balance relates to the impact this process would have on the solid waste disposal problem, both refuse and sewage sludge disposal. The production cost is indicative of the potential economic benefits that can be obtained from this system by producing a usable fuel gas.

Materials Balance for Process

A materials balance was performed on this process using a refuse of a composition shown in Table 28. Figure 20 shows the tons and volume of material flow through the process. This was developed in the following manner. Assuming a 100 ton/day raw refuse received, 78 tons of volatile solids plus 22 tons of inerts are

Table 28. Composition of a Typical Domestic Refuse

Component	% of Total by Weight	
	(Wet Weight)	(Dry Weight)
Paper	48.0	35.0
Leaves	9.0	5.0
Wood	2.0	1.5
Synthetics	2.0	2.0
Cloths	1.0	0.5
Garbage	16.0	8.0
Combustibles	78.0	52.0
Glass	6.0	6.0
Metal	8.0	8.0
Ashes, stone, etc.	8.0	6.0
Total moisture content	-	28.0

fed to the size reduction. The volatile solids contain 26 tons of water plus 52 tons of dry solids. An additional two tons of moisture is associated with the inerts. Assuming a density of 300 pounds per cubic yard as dumped from the collection vehicle, the volume of the refuse would be 667 cubic yards. The only change resulting from the size reduction would be to perhaps increase the density to 500 pounds per cubic yard with a volume of 400 cubic yards passing to the hydraulic separation unit.

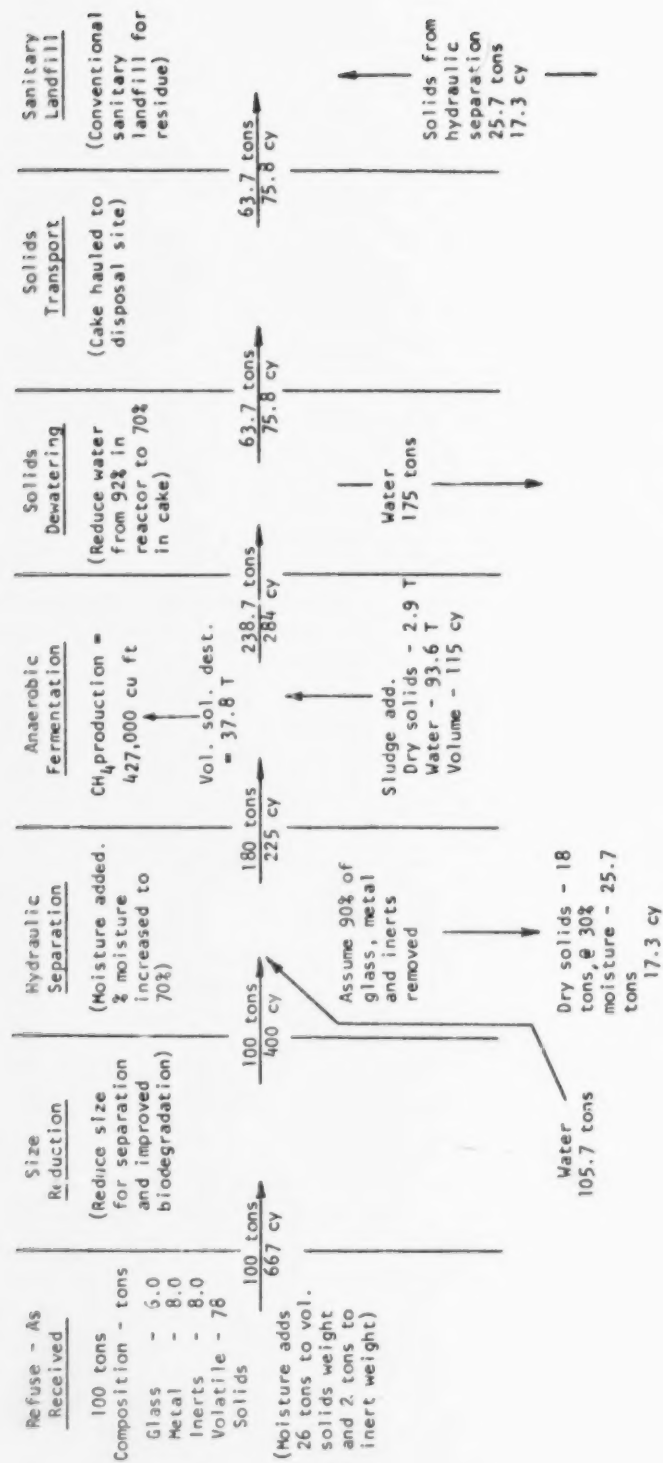


Figure 20. Mass-Volume Balance for Conversion of Refuse to Methane

Assuming the separation of the inerts is 90 percent efficient, then 18 tons of dry solids would be removed from the process stream at this point. The total weight would be dependent upon the amount of water retained with the solids. Laboratory measurements of the bulk density of this residue obtained from Cincinnati yielded a value of 2080 pounds per cubic yard. The specific gravity of the individual particles averaged 2.35. Because of the void space in the granular material, the retained water would probably not increase the volume. Therefore, the volume of this residue would be 17.3 cubic yards per 100 tons of refuse processed.

If the refuse discharged from the separation process has a moisture content of 70 percent, the water required in this step can be determined. A total of 28 tons is added with the original refuse. The 18 tons of inerts contain 7.7 tons of water at 30 percent moisture content. The 54 tons of solids discharged to the reactor will require 126 tons of water. To satisfy the water requirement, 105.7 tons of water would be added at this point.

The tonnage of wet refuse added to the reactor would be 180 tons per 100 tons of raw refuse processed. Assuming a specific gravity of 1.0, the volume of this feed would be 225 cubic yards. In addition to the refuse, sludge from the water pollution control plant would be added at this point. The volume of sludge would be 115 cubic yards consisting of 2.9 tons of dry solids plus 93.6 tons of water. The following calculations were used to develop the solids load to the reactor.

1. Population producing 100 tons of refuse/day at 7 lb/ca-day

$$100 \frac{\text{tons}}{\text{day}} \times \frac{2000 \text{ lbs}}{\text{ton}} \times \frac{1 \text{ ca-day}}{7 \text{ lbs}} = 28,500 \text{ population}$$

2. Primary sludge production--60 percent efficiency in primary sedimentation

$$0.22 \frac{\text{lb}}{\text{ca-day}} \times 28,500 \text{ ca} \times 0.6 = 3780 \frac{\text{lb}}{\text{day}} \text{ of total solids}$$

3. Secondary sludge production--30 percent BOD reduction in primary

$$0.2 \frac{\text{lb}}{\text{ca-day}} \times (1 - 0.3) = 0.14 \text{ lb } \frac{\text{BOD}}{\text{ca-day}} \text{ to secondary treatment}$$

Assume 0.5 lb of suspended solids produced per lb BOD₅ in an activated sludge process.

Pounds of suspended solids from secondary

$$0.14 \frac{\text{lb BOD}_5}{\text{ca-day}} \times 0.5 \frac{\text{lb S. S.}}{\text{lb BOD}_5} \times 28,500 \text{ ca} = 2000 \text{ lbs/day}$$

4. Total dry solids generated at water pollution control plant

$$3780 + 2000 = 5780 \text{ lbs/day at three percent solids}$$

$$\text{lbs wet sludge} = 5780/0.03 = 193,000 \text{ lbs}$$

5. Solids loading to the reactor

<u>Dry Solids</u>	<u>Water</u>
Refuse - 54 tons	Refuse - 126 tons
Sludge - 2.9 tons	Sludge - 93.6 tons
Total 56.9 tons	Total 219.6 tons

6. Solids content of reactor feed

$$\% \text{ Solids} = \frac{56.9}{(219.6 + 56.9)} = 20\%$$

At 80 percent moisture content, mixing of the reactor may require a significant amount of energy. However, the destruction of the volatile solids will reduce the solids content of the reactor substantially.

7. Volatile solids to reactor

Refuse	- 52 tons
Sludge - 2.9 tons at 70% volatile solids	- 2 tons
	54 tons

From the laboratory data, gas production was 4.4 scf per pound of volatile solids added at a 30-day retention time and at 35°C. This would correspond to a volatile solids reduction of about 40 percent. Operation at 60°C and 30-day retention time yielded a gas production of 7.2 scf per pound of volatile solids added with a corresponding solids reduction of 70 percent. Therefore, when operating at 60°C, 37.8 tons of organic solids would be converted to carbon dioxide and methane. Gas analysis showed this gas to contain 55 percent methane.

8. Methane production from volatile solids destruction at 60°C

$$54 \frac{\text{tons}}{\text{day}} \times 2000 \frac{\text{lb}}{\text{ton}} \times 7.2 \frac{\text{scf}}{\text{lb}} \times 0.55 = 427,000 \text{ scf}$$

The conversion of these solids to gas reduces the solids content in the reactor. Any mixing considerations must be based on the solids concentration in the reactor rather than on the concentration of solids in the feed slurry.

9. Solids concentration in the reactor - 30 day retention time

	Tons of Solids Remaining Per Day
Volatile solids = 54 - 37.8 =	16.2 tons
Fixed solids = 56.9 - 54 =	2.9 tons
Total =	19.1 tons/day
Tons of water added per day =	219.6
Solids content in digester =	$\frac{19.1}{(219.6 + 19.1)} \times 100 = 8\%$

The effluent from the digester consists of 219.6 tons per day of water plus 19.1 tons per day of solids or a total of 238.7 tons per day at a moisture content of eight percent. The specific gravity of this slurry would be approximately 1.0. Therefore, the volume of slurry to dewater would be 284 cubic yards per day.

Prior to landfilling, this slurry must be dewatered. Laboratory filter test-leaf studies suggest that a moisture content of 70 percent is achieved without the use of chemicals to condition the sludge. At 70 percent moisture, the 19.1 tons of dry solids would produce a cake weighing 63.7 tons per day. The water, 175 tons, would have to be processed for disposal, discharged to the water pollution control plant, or up to 105.7 tons could be recycled to the process.

Because the specific gravity of the solids in the dewatered slurry would be about 1.1 to 1.3, the specific gravity of the dewatered material would be approximately 1.0. Therefore, the volume of this material for landfilling would be 75.8 cubic yards. However, a reduction of the moisture content to 60 percent would leave a final volume for disposal of only 56 cubic yards having a weight of 47.7 tons. Adding to this the 17.3 cubic yards of inert residue from the hydraulic separation step, the volume for final disposal would be between 74 and 81 cubic yards per 100 tons of original refuse. This contrasts to the perhaps 667 cubic yards of loose refuse without any processing. The degree of compaction that can be achieved by landfilling the process residue is not known. Assuming a 20 percent reduction in volume due to compaction, the volume of the compacted process residue would be between 59 and 65 cubic yards per 100 tons of refuse processed. This can be compared to the volume required by the unprocessed refuse. Assuming compaction of the refuse produces a density of 1000 pounds per cubic yard, the 100 tons of refuse plus digester residue from 2.9 tons of sewage solids would require 206 cubic yards of fill space. Therefore, only 31.6 percent of the landfill volume is required after processing.

Unit Processes - Shredding

Size reduction is a well known process, having been applied extensively in many industrial operations. In recent years, some of these units have been modified for use in reducing the particle size of the refuse prior to processing by various means. Drobny *et al.* (1971) evaluated the cost associated with the processing of municipal refuse. The power requirements for several installations are shown in Figure 21. A probable power requirement for shredding of municipal refuse is approximately 10 HP-hr/ton. The power requirement is a function of particle size. The following product size factors, developed from experimental data, show the increased power required for smaller size particles. The size of product required will shift the power requirements from near 10 HP-hr/ton for a six-inch size particle to approximately 25 HP-hr/ton for a one-inch size particle.

<u>Particle Size</u>	<u>Product Size Factors</u>
6 inch	1.00
4 inch	1.39
2 inch	1.64
1 inch	2.38

Actual cost data for a Gondard Mill pilot operation is shown in Table 29 (Porter 1970). This shredding process was used in conjunction with the Madison, Wisconsin landfill project. The author states that these costs data are high because certain segments of the plant were over designed and the refuse feed and product removal equipment were not sized to use the mill at capacity. Additional labor was required because of inadequacies in the design. Also, the mill was only operated for 5.33 hours per day rather than seven hours at a capacity of nine tons per hour. Adjustment of the costs for a mill operating for a 7-hour shift per day at capacity reduced the cost to \$5.05 per ton for the 5 inch size particle. The major cost reduction was in labor, being reduced from \$3.46 per ton to \$1.89 per ton. Incremental cost decreases occurred in all categories except hammer wear and mill maintenance. Major cost factors were amortization and labor for this system.

Additional costs reduction in the pilot plant cost can be expected. Installation of a Tollemache Mill at this pilot plant produced a shredding cost of \$4.92 per ton. This cost is based upon operation of the mill only 5.3 hours per day rather than seven hours.

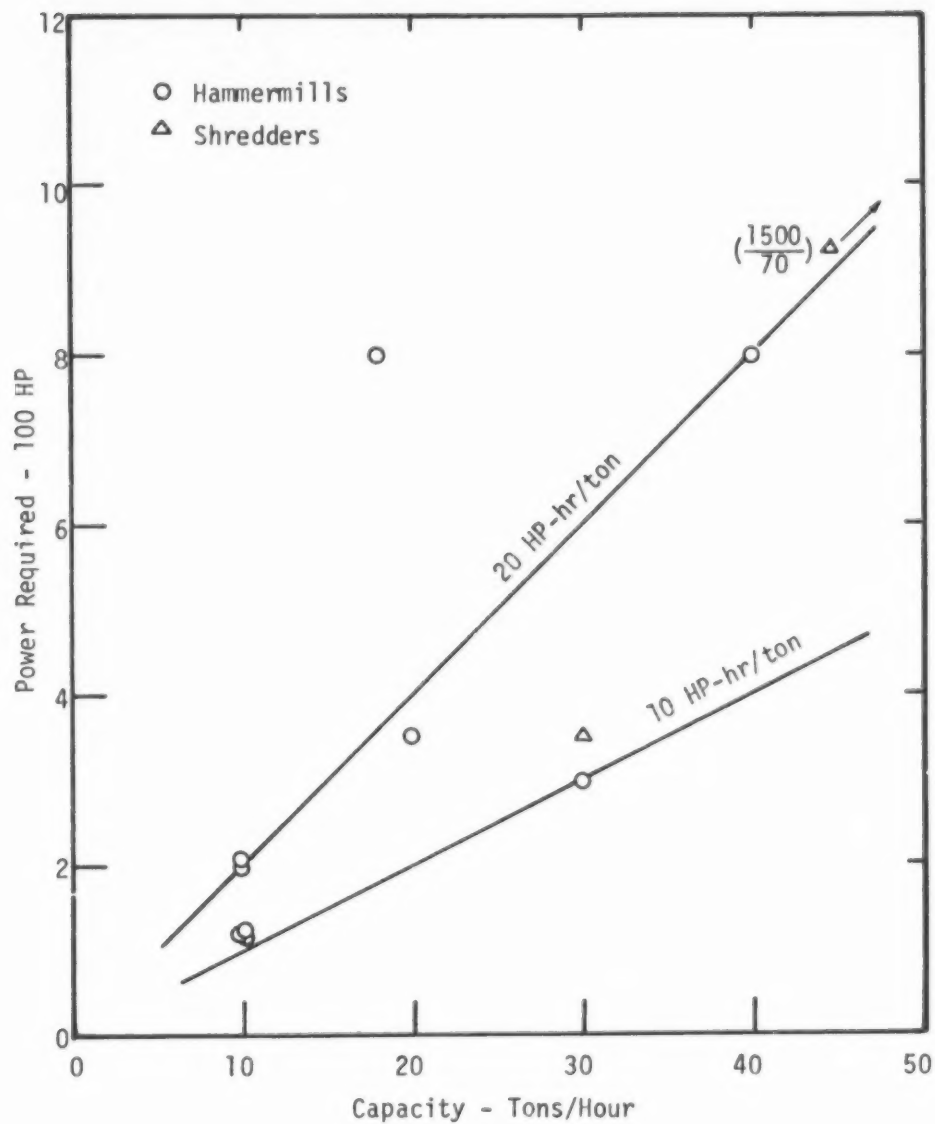


Figure 21. Power Requirements for Hammermills and Shredders (Drobney *et al.*, 1971)

Table 29. Shredding Cost for 5 Inch Grate Size (Gondard Mill)
(Period of Operation--June 1968 - May 1969)

Cost Item	\$/Ton	% Total Cost
Labor	3.46	45.2
Amortization	2.80	37.4
Power	0.30	4.0
Utilities	0.32	4.3
Hammer wear	0.15	2.0
Mill maintenance	0.08	1.1
General supplies	0.17	2.3
Other	0.19	2.5
Total	7.47	100.0

A prediction of cost for one and/or two shift operation and different number of mills is summerized in Table 30 (Ham *et al.*, 1971). In this calculation, the size reduction equipment is hammermills with 15 tons per hour capacity each. They are to be operated seven hours per shift and 245 days per year. Where one man can monitor two mills, the cost ranged from \$2.42 per ton for a minimum number of mills and one shift work to \$1.21 per ton for a plant working two shifts, with capacity of about 200,000 tons per year.

In a similar analysis, Porter (1970) projected the cost associated with the Gondard Mill based on the number of mills and one or two shift operation. The predicted cost is presented in Table 31. The cost in this calculation is \$1.96 to \$3.95 per ton. These values are highly dependent on operational conditions and characteristics of the system. Applying these analysis to another system requires careful study. However, improvement in the system is expected in the future and the cost may be reduced. Also operating the system at high capacity and making full time use of the mills provides less cost. A value of \$1.5 to \$2.5 per ton seems to be relatively reasonable for a cost analysis.

Unit Processes - Separation

Most of the processes involving recovery of solid waste components require a separation step. Many different systems have been developed for various purposes, particularly for industrial application. The efficiency with which separation is affected and the degree to which the desirable materials can be segregated

Table 30. Annual Costs of Combinations of Mills and Work Shifts
(Tollemache Mill)

	Number of Mills Operated			
	1	2	3	4
<u>One shift</u>				
Daily tonnage	105	210	315	420
Milling plant				
operation cost	40,500	59,500	72,200	91,900
Depreciation	21,600	38,000	50,900	64,200
Total cost	62,100	97,500	123,100	156,100
Annual tonnage	25,700	51,400	77,100	102,800
Cost, \$/ton	2.42	1.89	1.60	1.57
<u>Two shifts</u>				
Daily tonnage	210	420	630	840
Milling plant				
operation cost	76,000	114,700	139,200	178,100
Depreciation	24,300	41,300	55,900	70,400
Total cost	100,300	156,000	195,100	248,500
Annual tonnage	51,400	102,800	154,200	205,600
Cost, \$/ton	1.95	1.52	1.26	1.21

Table 31. Annual Cost of Combinations of Mills and Work Shifts

	Number of Mills			
	1	2	3	4
<u>One shift work</u>				
Total cost, \$/yr	60,900	95,800	120,500	152,700
Annual tonnage	15,400	30,900	46,400	61,800
Cost, \$/ton	3.95	3.17	2.60	2.48
<u>Two shift work</u>				
Total cost, \$/yr	100,300	153,400	190,600	242,500
Annual tonnage	30,900	61,800	92,700	123,600
Cost, \$/ton	3.24	2.48	2.05	1.96

are significant factors contributing to the cost and marketability of the recovered product. Until recent years hand sorting was the major means of separation when the objective was basically for salvage purposes. Hand sorting is no longer a feasible technique for separation because of low salvage price, low efficiency of the system, labor costs and human fallibility. Drobny *et al.* (1971) report a cost of \$3.50 to \$5.00 per ton of newsprint separated when using low cost labor.

There is inadequate information to evaluate the effectiveness of mechanical separation processes for solid waste. Experimental data need to be collected to determine interrelationships among such properties as particle size and distribution in the mixed feed, separation efficiency, capacity, moisture content and material type. A complete separation of all components in the refuse is seldom possible. Mechanical separation is used to divide a mixture into two or more fractions where each group is uniform in a certain property. Solids can be graded based on differences in dimensions, shape or specific gravity. In special cases, the roughness of the surface, color or magnetic properties of the particle can be employed. In general, two types of separation techniques can be used, i.e. wet or dry process. For specific cases, an investigation should be undertaken to select the optimum separation technique. Wet techniques are not used prior to a dry recovery process because of the dewatering and drying costs involved.

The characteristic of each separation unit process can be found elsewhere (Rietema and Verver, 1961). However, there is a universal lack of cost and operational data. An important part of the cost involved is auxiliary items. Hence the sum of the unit costs may not exactly represent the total cost. Also separation and size reduction should be considered together from the particle size view point. It is clear that different materials in mixed solid waste are reduced to different sizes by the shredding step. Screening may present better efficiency for separation of some components, particularly for such items as glass and grit.

Using a digester for anaerobic fermentation of refuse with characteristics as shown in Table 28 requires the separation of inert materials. The inert materials account for about 22 percent by weight of the solid waste. The primary disadvantages of these materials are the volume they occupy in the reactor and the added materials handling problems. The abrasiveness of the inert material also will cause excessive mechanical wear on the equipment. These materials may also be the source of toxic substances and decrease the organisms' activities. The initial investigations in this study indicated that the biological mechanism was not inhibited by using unseparated refuse feed. However, the

accumulation of potentially toxic materials in the reactors should be avoided.

With the exception of plastics, the specific gravity of the inert materials is high. This property can be advantageously employed in the separation process. Hence for similar size particles, an air or hydraulic separation process offers a feasible technique for separation of the various components. Also, the use of screens prior to the separator may introduce a more efficient system by removing much of the fine grit and glass particles. For gravity separation, water is usually preferred to air because air classification requires a more narrow size distribution range. Also, particle shape has a more significant effect on the air separation technique. Additional problems such as a uniform airflow make this separation technique difficult.

The problem with hydraulic separation is the feed concentration. Because most of the techniques are applicable for low slurry concentrations, a thickening or dewatering step is required prior to the reactor in order to take advantages of the concentrated feed for achieving significant retention times with small reactor volumes.

Based on the degree to which the desirable materials can be removed, some technique such as a "stoner" can be employed. The power required is about 1 HP per ton and the amortized capital cost varies \$0.10 to \$0.15 per ton (Drobny *et al.*, 1971). If the recovery of inert material is required, a more efficient system should be introduced. The cost of magnetic separation is about \$13.5 per ton of material recovered or about \$0.90 per ton of refuse. Ballistic separation also provides a classification into two groups, organic and inorganic materials. However, more research and pilot plant study are required for the evaluation of the separation system. For this initial investigation, a cost of about \$0.50 per ton may be assumed. This is in the middle of the reported costs, but these costs are highly dependent upon the type of system and separation efficiency required.

Unit Process - Anaerobic Fermentation

From economic considerations, the most important variables in the anaerobic process are temperature and detention time. The other factors influencing the anaerobic process, such as pH, are assumed to be controlled without any economic impact on the process. These two factors were assumed to be the more significant variables and the optimum conditions are defined in terms of temperature and detention time. The laboratory data obtained in this study were used as input to the model.

Gas production for various temperatures and detention times is shown in Figures 12 and 16. Table 32 was developed from these figures. The variation between two consecutive points is taken as a linear function. The loading rate for which these data were obtained was 0.21 pounds of dry solids per cubic foot per day. The gas production was found to be 11 scf per pound of volatile solids destroyed. This was the basis for construction of Table 33.

Table 32. Gas Production (Cu Ft/Lb Dry Solids)

Temperature (°F)	Retention Time (Days)					
	4	8	10	15	20	30
95.0	1.40	2.30	2.58	2.86	2.95	3.05
104.0	2.33	2.98	3.20	3.60	3.85	4.10
105.8	2.49	3.09	3.25	3.71	3.95	4.20
107.8	2.59	3.14	3.33	3.74	3.95	4.20
109.4	2.58	3.09	3.26	3.67	3.82	4.09
113.0	2.08	2.55	2.70	3.05	3.14	3.36
118.4	2.64	3.14	3.33	3.57	3.82	4.06
127.4	3.33	3.88	4.07	4.29	4.51	4.70
132.8	3.74	4.20	4.35	4.54	4.75	4.88
140.0	4.23	4.49	4.57	4.79	4.93	4.96

Table 33. Volatile Solids Destruction (Percent)

Temperature (°F)	Retention Time (Days)					
	4	8	10	15	20	30
95.0	17.4	28.6	32.1	36.6	36.7	38.0
104.0	29.0	37.1	39.8	44.8	47.9	51.0
105.8	31.0	38.5	40.5	46.2	49.2	52.3
107.8	32.2	39.1	41.5	46.6	49.2	52.3
109.4	32.1	38.5	40.6	45.7	47.6	50.9
113.0	25.9	31.7	33.6	38.0	39.1	41.8
118.4	32.9	39.1	41.5	44.4	47.6	50.6
127.4	41.5	48.3	50.7	53.4	56.2	58.5
132.8	46.6	52.3	54.2	56.5	59.1	60.8
140.0	52.7	55.9	56.9	59.6	61.4	61.8

No solids recycling was employed and a complete mixed system was assumed where the outflow solids concentration is equal to the solids concentration in the reactor. This assumption can be employed to calculate the slurry content of the outflow based on the volatile solids destroyed in the reactor.

A computer simulation technique was employed to calculate the cost of producing gas. The total cost of the system is divided into two categories, capital and operational cost. The following discussion presents the cost analysis of the important factors which contribute to the total cost.

Capital Cost

The cost of the reactor is an important factor in the cost analysis. This cost relationship can be expressed in terms of the volume of the reactor. The volume of the reactor based on fixed feed concentration is equal to:

$$V = \frac{M}{C\rho} t \quad (14)$$

where: ρ = density of feed stream, lb/cu ft
 V = volume of reactor, cu ft
 M = lbs of solid refuse fed per day
 C = feed concentration, %
 t = detention time, days

If a constant loading rate is used, then C in Equation 14 can be presented as

$$C = (F)(t)(V) \quad (15)$$

where: F = loading factor, lbs/cf/day

However, the model permits both kinds of reactor volume design, i.e. a fixed or variable loading factor.

The cost formula developed by Smith (Eckenfelder and Adams, 1972) was selected from the existing formulae and a construction index value of 1.23 was used to update the cost into January 1973. This index was obtained from sewage treatment cost index (*Engineering News Record*, 1972). Use of the ENR Construction Cost Index introduces a higher value than the sewage treatment cost index.

$$CC = V(1.65 + 17/V^{0.87}) \quad (16)$$

where: CC = total capital cost in \$1000, January 1973
V = reactor volume in 1000 cf

A life of 25 years and a current value of interest rate of six percent were used. A factor of 0.07823 for amortization of the capital cost corresponds to this project life and interest rate.

Operating Costs

The operating costs are separated according to the different functions. The most important operating costs are heating, mixing and residue dewatering.

Heating. Heating costs are an important design factor. Elevating the temperature to increase the rate of solids destruction requires more heat energy. However, a reduction in the heating requirements can be obtained by operating at a shorter detention time. This, in turn, leads to a lower capital cost. The heat must be adequate to increase the feed stream temperature to the reactor temperature level and also cover the heat loss from the system.

The heat required to elevate the temperature of the feed stream is proportional to the mass flow rate and the difference between the feed stream and the reactor temperature. However, a part of this heat can be recovered from the discharged flow by use of a counter current heat recovery system. An extreme condition for an initial temperature of 35°F for the feed slurry was selected for the economic analysis.

$$HI = S(c)(T_1 - T_0) \quad (17)$$

where: S = feed stream, lbs/hr
c = specific heat constant, BTU/lb-°F
T₁ = reactor temperature, °F
T₀ = feed stream temperature, °F
HI = heat required, BTU/hr

An important part of the heat loss results from the surface heat loss from the reactor. A cylindrical digester with a height equal to the diameter was assumed in order to achieve the minimum surface

area. Equation 18 is used to calculate the heat loss from the surface area.

$$Q = \sum U_i A_i (T_1 - T_i) \quad (18)$$

where: Q = heat loss, BTU/hr
 U_i = heat loss per unit area per hour, BTU/hr-ft²-°F
 T_i = outside temperature, °F
 A_i = surface area, sq ft

Operating the digester at high temperature introduces heat loss by evaporation of the water which leaves the system as a gaseous product. This heat requirement can be calculated from Equation 19.

$$Q = \lambda w \quad (19)$$

where: λ = unit latent heat, BTU/lb
 w = weight of the material changing state, lb
 Q = heat loss, BTU

The actual volume of a gas generated by water vapor is obtained from Equation 20.

$$V_o = 16.4 V \left(\frac{P - P_w}{460 + T_1} \right) \quad (20)$$

where: P_w = vapor pressure of water at T_1
 P = atmospheric pressure, 29.6 in. Hg
 $V - V_o$ = volume of the water vapor

An empirical relation was developed to present the heat loss versus gas production. Equation 21 shows this relationship.

$$Q_L = 1 + 0.16 (T_1 - 32) \quad (21)$$

where: T_1 = reactor operating temperature, °F
 Q_L = heat loss, BTU/cf gas produced

The metabolism of organic compounds releases heat energy. The heat released in aerobic systems is substantial and raises the temperature of the system. However, in anaerobic systems the metabolic heat released is low at about 1.6 BTU/lb refuse. Most of energy released from the organic material is in the form of methane gas which can be recovered by using it as a fuel.

Mixing Energy. Mixing is a relatively important operational factor. The criterion for mixing is to keep the organic materials in suspension in order to increase the contact between the substrate and microorganisms and improve the fermentation process. Mixing should be adequate to prevent deposition of organic materials and to keep the reactor contents relatively homogenous. The energy required for mixing is a function of the volume, solids concentration, the type of the solid material and the nature of the mixing equipment. Unfortunately, limited data are available on mixing requirements for anaerobic processing of concentrated organic refuse. Studies of high rate municipal waste digestion reported that a circulating velocity of 1.5 to 2 fps at the floor is necessary in order to prevent any substantial amount of deposition of material. The minimum velocity required is dependent upon the characteristics of the particles. Minimum velocities as high as 2.5 fps have been reported (WPCF, 1966). When operating at high slurry concentrations, continuous mixing is important because it is sometimes difficult to resuspend the deposited materials.

There are several mixing techniques that have been used for anaerobic fermentation processes. Sludge recirculation, gas circulation, pumping and mechanical draft-tube mixing are the most common. Each process has some advantages and disadvantages. In order to determine the proper mixing conditions and the most economical mixing equipment, more study seems to be required. Also, the type of the size reduction and separation processes may affect the mixing requirements. Based upon current practice for digester mixing, a value of 0.185 HP per 1000 cubic feet of reactor volume was used for four percent slurry concentration (McMahon, 1965). The following equation was used to calculate mixing power required for different slurry concentrations.

$$P = 0.185 \times \frac{C}{4} \times V \quad (22)$$

where: P = power requirements, HP
C = reactor slurry concentration, %
V = reactor volume, 1000 cu ft

Reactor Residue Dewatering. The dewatering operation is related to the slurry concentration and the characteristics of the material resulting from the anaerobic fermentation of the refuse. For this reason, operating costs for dewatering are taken as part of the total operating cost. The filter yield was assumed to be linearly proportional to the slurry concentration. The power requirement was estimated to be 1/8 HP per square foot of filter area. The laboratory filterability studies found higher filter yields from the concentrated slurry. Therefore, this model requires more filter area for the lower slurry concentrations which result from operation of the system at high temperatures and long retention times. However, the mass of solids available for dewatering is reduced under these conditions.

Additional studies of the dewatering step are necessary. When operating with a relatively low slurry concentration, a thickening step prior to dewatering of the residue may be advantageous. However, for the purpose of this analysis, the dewatering costs were based upon the characteristics of the reactor effluent. The filter area is the controlling factor in determining the capital cost of the filter. Equation 23 shows the relationship used to determine the capital cost for the filter. The area requirements were determined from the reactor effluent slurry concentration using filter loadings found from the laboratory studies.

$$\text{Log } C = 0.65 - 0.66A \quad (23)$$

where: C = cost, \$100 per square foot of filter area
A = filter area, 100 square feet

Economic Evaluation of the Anaerobic Fermentation Process

A computer simulation technique was employed in order to evaluate the process and to find the least cost condition for gas production. As previously mentioned, temperature and detention time were taken as major variables. The program for this simulation is shown in Appendix D. The input data are given on the first page of Appendix D.

The concentration of the reactor slurry is controlled by the concentration of the feed slurry and the quantity of solids converted to gas. Table 39 in Appendix E shows the calculated slurry concentration in the reactor at various feed concentrations, temperatures and retention times. As would be expected, the minimum solids concentration in the reactor for a given feed slurry concentration is achieved at the highest temperature and longest detention time. The effect of temperature and retention time on the reactor slurry concentration is graphically displayed in

Figure 22. These data represent only those conditions resulting from a feed slurry concentration of 20 percent solids. This figure shows the effect of the efficiency of converting the volatile solids to gas.

The energy required for mixing is a function of reactor volume and reactor slurry concentration. The mixing costs shown in Table 40, Appendix E, are calculated on the basis of the reactor slurry concentration and the reactor volume determined by the retention time. The maximum mixing energy required is at the lowest temperature and longest retention time. The cost of power for mixing ranges from \$2,350 per year at 140°F and 4.0 days retention time to \$20,700 per year at 95°F and 30 days retention time. These calculations are based on a system sized to process 100 tons per day of dry refuse. Based on the refuse shown in Table 28, this would be equivalent to 139 tons per day of refuse as received.

The heat requirements are primarily controlled by the reactor operating temperature. Retention time, as it relates to reactor size, also influences the heat requirements. The heat required to maintain the temperature under various conditions is given in Table 41, Appendix E. These calculations are based on a 75 percent recovery of the heat in the reactor effluent slurry. It is obvious that the minimum heat requirements are at the lowest temperature and shortest detention time. Increasing feed concentrations decrease the heat requirements for two reasons. Smaller reactors are required for a given refuse quantity and retention time resulting in less heat loss from the reactor surfaces. Also, increased feed slurry concentrations reduce the mass of water that must be heated when passing through the system. The heat requirements for a 100 ton per day (dry refuse) system operating with a feed slurry concentrations of 6 to 20 percent solids are shown in Figure 23. The heat requirements are shown for a system operating at the four day retention time. The feed slurry concentration is the major factor in heat losses. The dashed line shows the heat required to maintain the temperature if the retention time is increased from four to eight days when the feed slurry concentration is 20 percent solids.

The dewatering costs are primarily controlled by the feed slurry concentration. For a given slurry concentration, the change in reactor slurry concentration at different retention times and temperatures is small in comparison to the concentration variation range resulting from the substantial variations in feed slurry concentrations studied. Table 42, Appendix E shows the dewatering costs for a 20 percent feed slurry. The effect of the feed slurry concentration is shown in Figure 24. In this figure, the solid lines represent the dewatering costs for the 100 ton (dry refuse) per day system operating at four days retention time.

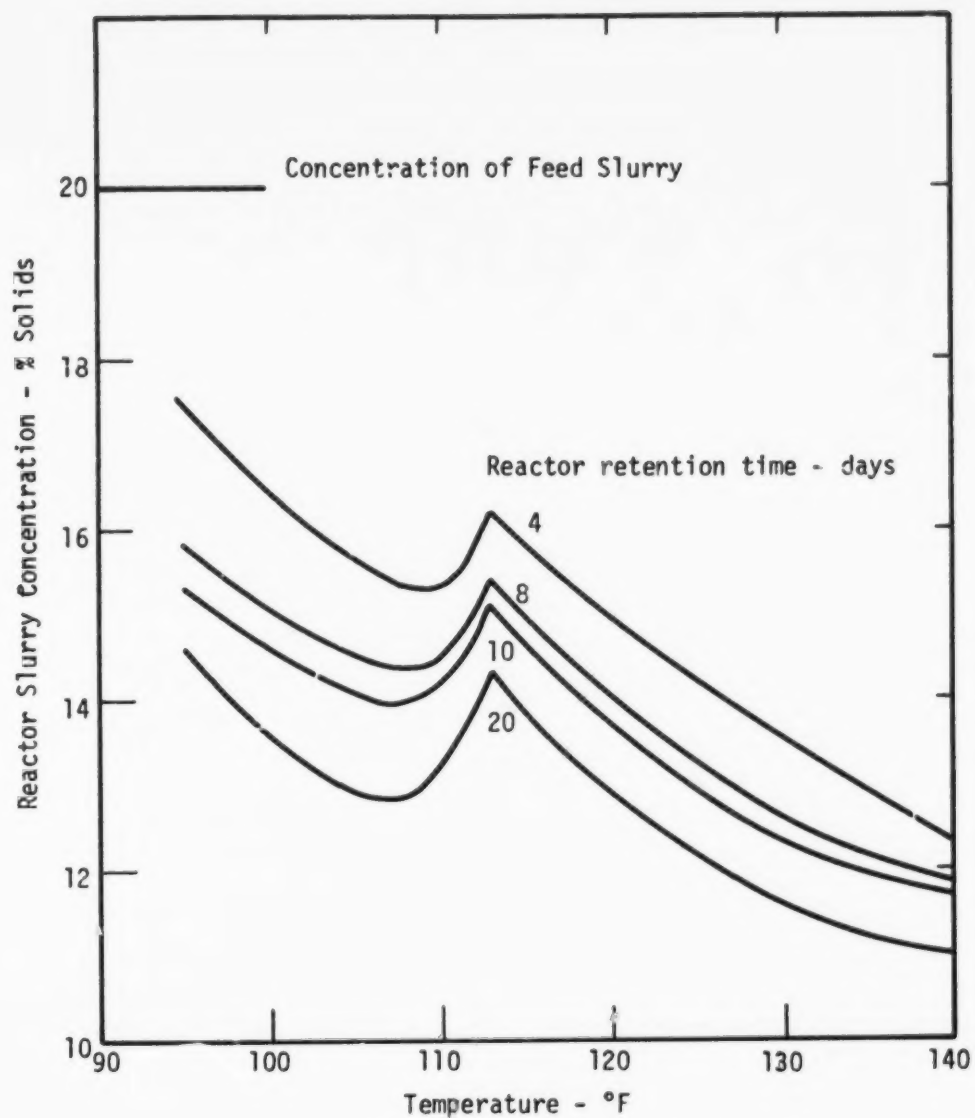


Figure 22. The Effect of Temperature and Retention Time on Reactor Slurry Concentration

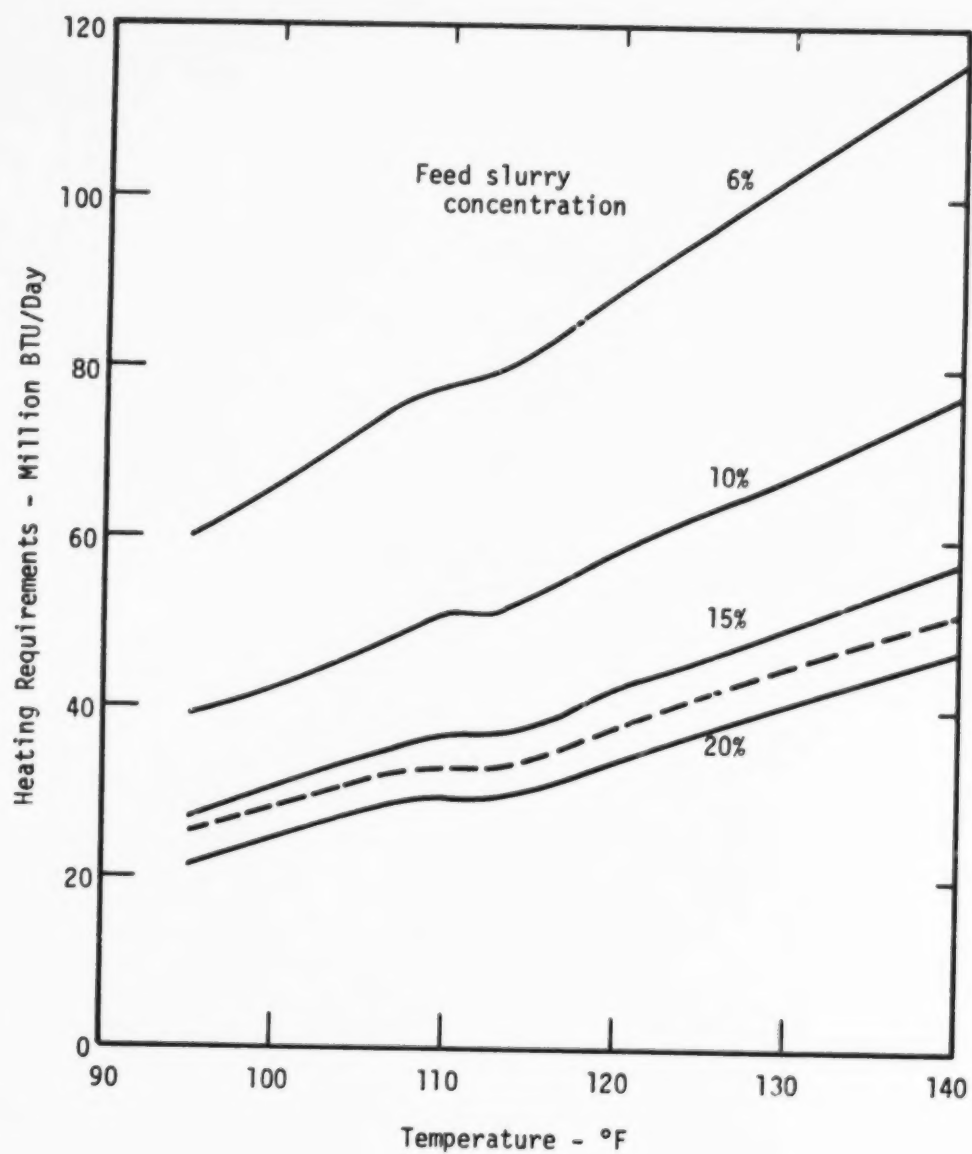


Figure 23. Heating Requirements for Operating at Various Temperatures and Feed Slurry Concentrations

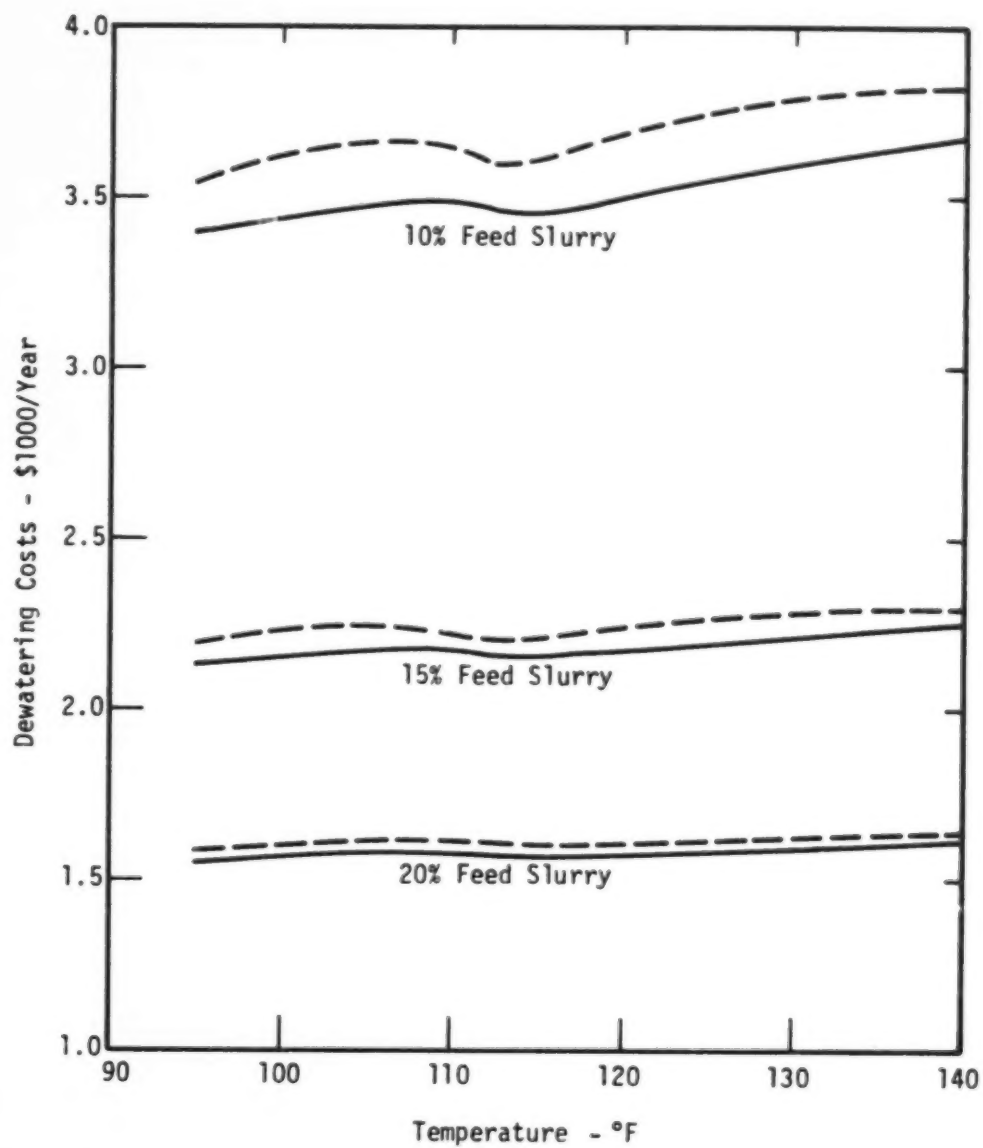


Figure 24. Residue Dewatering Costs Under Various Operating Conditions

The dashed lines represent the system operating at a 30 day retention time. The dewatering costs are slightly higher for the longer retention times due to the more dilute slurry. The calculated filter yield is lower for the more dilute slurry. Since the same volume of water must be filtered, the cost increases slightly for the longer retention times. This figure shows that dewatering costs are primarily a function of feed slurry concentration. Temperature and retention time have very little effect on this cost item.

The costs discussed above can be combined to determine the operating and total costs for the anaerobic fermentation of refuse. These do not include the preparation costs associated with shredding and separation nor disposal costs for the dewatered residue. Table 34 presents a breakdown of these cost factors. It is evident that the capital cost is a significant cost contributor under all conditions.

Table 34. Operation and Capital Cost - Plant Capacity of 100 Tons per Day of Dry Refuse - 20 Percent Field Slurry

Item	30 days, 95°F		4 days, 140°F	
	Annual Cost \$1000	Percentage	Annual Cost \$1000	Percentage
Heating	6.59	7.0	9.38	39.3
Mixing	20.70	22.0	2.35	9.8
Dewatering	1.58	1.7	1.61	6.7
Total operation	28.87	30.7	13.34	55.8
Capital cost	64.90	69.3	10.50	44.2
Total	93.77	100.0	23.84	100.0

Operating at the minimum temperature (95°F) and maximum retention time generates a total annual cost of \$92,770 for a system receiving a feed slurry with 20 percent solids. The capital costs account for 69.3 percent of these costs. For these conditions, mixing costs are also high, accounting for 22.3 percent of the total costs. Operation of the system at the maximum temperature and minimum retention time shifts the costs downward. The capital cost contribution is reduced to 44.2 percent of the \$23,840 annual cost. The mixing cost is considerably reduced. At this high temperature, the heating cost assumes a much higher proportion of

the cost, accounting for 39.3 percent of the total annual costs. The total annual costs for all operating conditions investigated are given in Table 43, Appendix E.

The total annual costs become meaningful only when they are compared to the benefits. The net benefits are calculated from the quantity of gas produced. The value of the gas produced was calculated on the basis of selling the gas for \$0.45 per 1000 cu ft. Based upon a heating value of 600 BTU per cubic foot, the energy would sell for \$0.75 per million BTU. The maximum benefits accrue under conditions that yield the maximum gas production. The temperature and retention time are the controlling factors in this analysis. The net benefits have the production costs subtracted from the value of the gas produced. The net benefits for all conditions analyzed are given in Table 44, Appendix E. Figure 25 is a plot of the net benefits against temperature for various retention times. This plot is for a feed slurry concentration of 20 percent solids. The maximum net benefits are obtained when the system operates at the minimum retention time and maximum temperature. Examination of this figure shows the maximum benefits are obtained at 140°F and a retention time of four days. In the mesophilic temperature range, the maximum benefits are obtained at 108°F with a retention time of eight days.

The effect of retention time and feed slurry concentration on the net benefits at the optimum thermophilic and mesophilic temperature can be seen in Figure 26. Slurry concentration is a major factor in determining the net benefits under all operating conditions. When operating at 108°F, the retention time is not a major factor in net benefits when operating with a feed slurry concentration of 15 to 20 percent. A retention time between 4 and 20 days yields essentially the same net benefits. The cost of residue disposal is not included. Therefore, the total system net benefits would probably be at the 20 day retention time. This will yield less residue for disposal. Other factors such as operating stability would suggest operation at the longer retention times. Since one of the major purposes of this process is methane production, it would be logical to operate under those conditions that yield the maximum gas without significantly reducing the net benefits.

In the thermophilic range, the curves for 15 and 20 percent feed slurry concentrations are not as flat. However, the change in net benefits between four and ten days retention time is small. When the cost of residue disposal is included, the optimum retention time may well shift to the longer times. As with mesophilic temperature, operating stability and maximum gas production may warrant operating in the range of 10 days retention time. Since the cost of pH control has not been included in this analysis, the lower caustic requirements

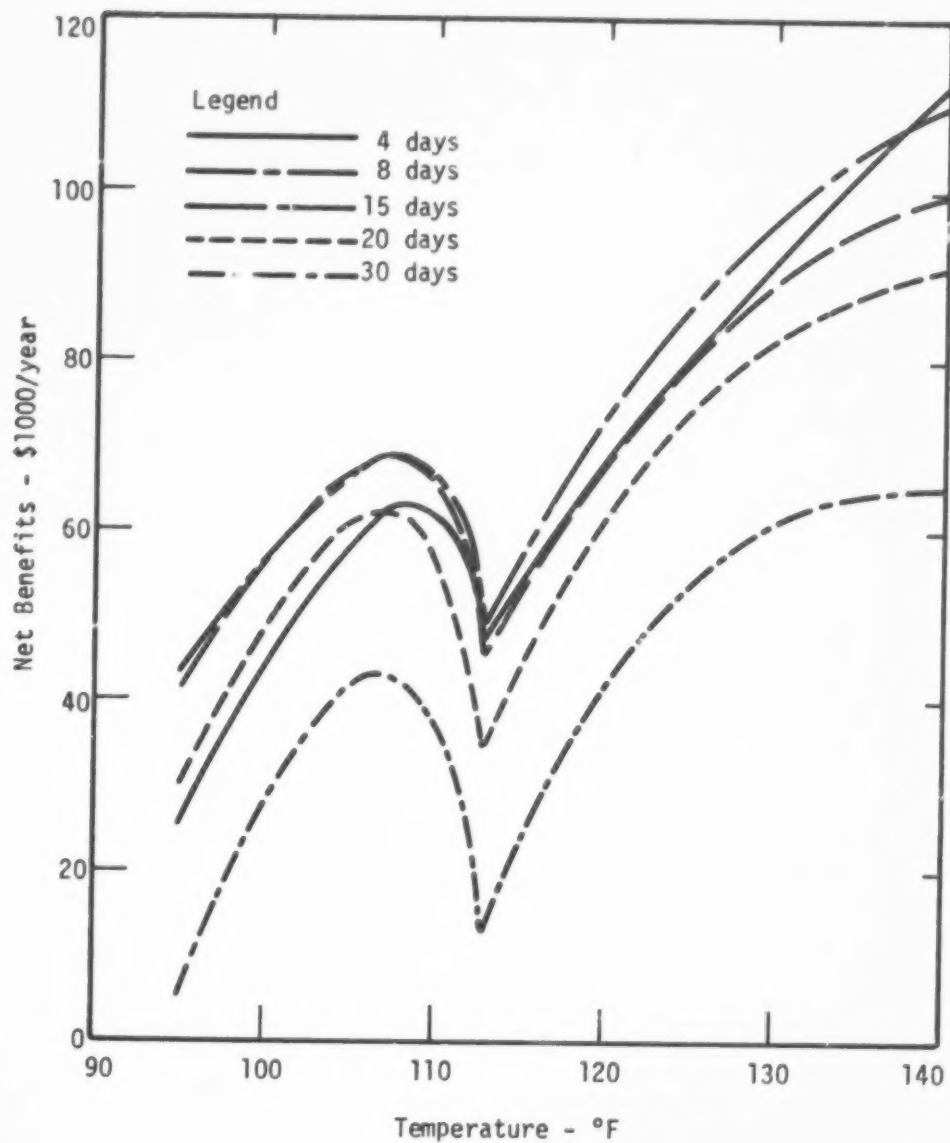


Figure 25. Net Benefits Resulting at Various Temperatures and Retention Times, Feed Slurry - 20% Solids

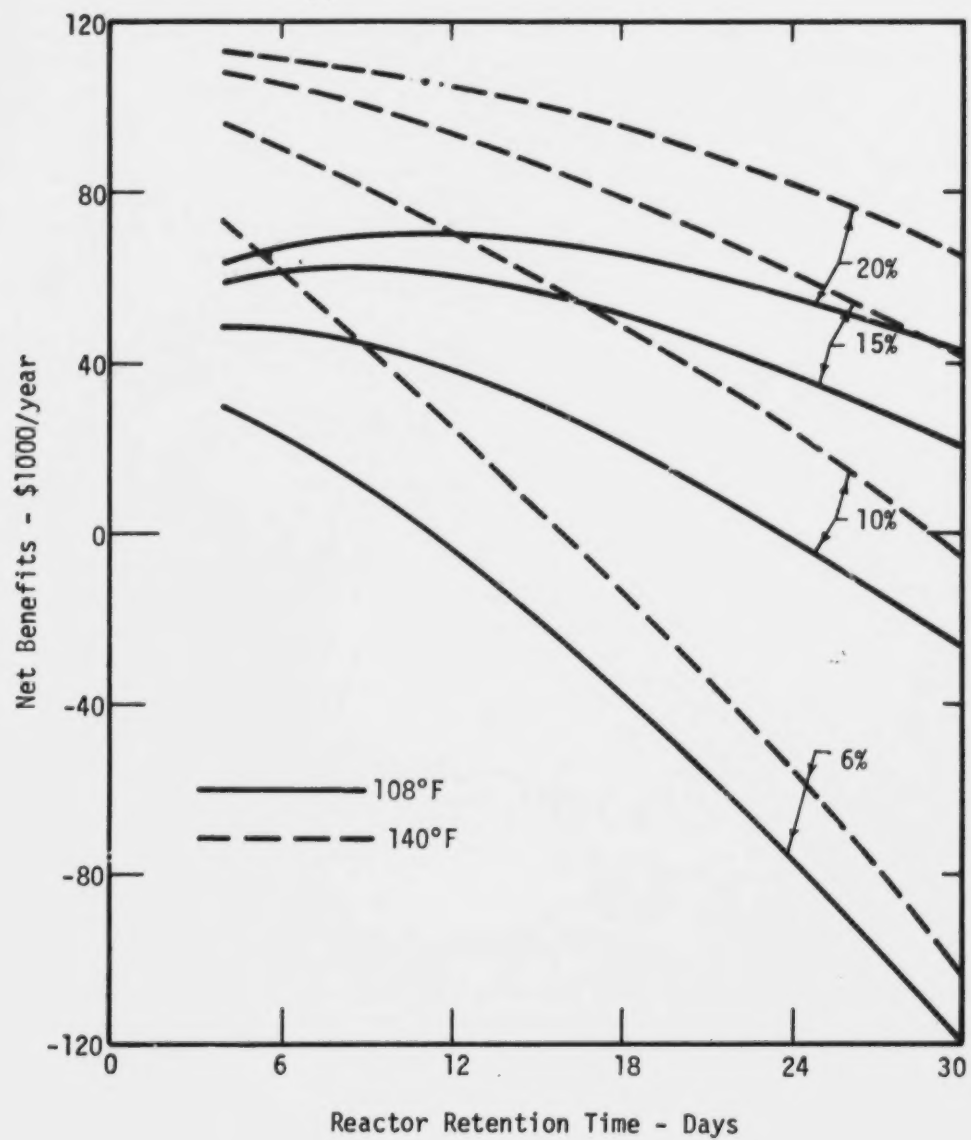


Figure 26. Net Benefits at Optimum Temperature for Various Retention Times and Feed Slurry Concentrations

at the longer retention time may also shift the net benefits slightly to the longer retention period.

Although the net benefits associated with thermophilic operation are about 50 percent greater than when operating in the mesophilic temperature, it is not clear that the thermophilic range will be optimum. Operating at the higher temperatures may require an increased operation and maintenance cost. Also, these costs assume 75 percent recovery of the heat content of the reactor effluent. Recovery of only 25 percent of the heat contained in the reactor effluent increases the heat required by about 60 percent. This can contribute a significant cost increase under certain operating conditions. With a feed slurry of 20 percent, the heating costs would increase \$5,600 per year at 140°F operating temperature. This would reduce the net benefits associated with the thermophilic temperature range. Many of these factors must be investigated on a large scale system to determine the exact optimum operating conditions.

Mixing costs associated with the digestion of sewage sludge have been applied to this system. Increased power requirements for mixing could be significant at the longer retention times. At a 30-day retention time, mixing accounts for about one-fifth of the total costs. Increasing the power requirements from 0.185 HP per 1000 cu ft to 1.0 HP per 1000 cu ft for a four percent reactor slurry would increase the total annual cost from \$93,770 to \$185,000 for the conditions shown in Table 34. A similar analysis of the four day retention time shows that the total annual costs increase from \$23,800 to \$34,190. This is a 43 percent increase as compared to a 100 percent increase at the 30-day retention time. Conversely, a reduction in power requirements at the higher slurry concentration could significantly reduce the mixing costs. Therefore, due consideration must be given to obtaining better data on the mixing power requirements.

The cost of producing gas is shown in Figure 27. This figure shows the minimum cost per unit of gas produced for each temperature and feed slurry concentration. This curve shows the minimum cost of gas production is at 140°F. Table 45, Appendix E, presents the gas cost for all operating conditions. For the 15 percent feed slurry concentration, this cost was 9.6 cents per 1000 cu ft of gas. In the mesophilic temperature, the minimum cost was 13.7 cents per 1000 cu ft. At a feed slurry concentration of 20 percent, the mesophilic and thermophilic costs were 11.1 and 7.8 cents per 1000 cu ft respectively. Increasing the feed concentration to 25 percent decreases the cost of the gas to 9.5 and 6.7 cents per 1000 cu ft. As previously indicated, the power required for mixing may significantly alter the operating conditions that yield the maximum net benefits.

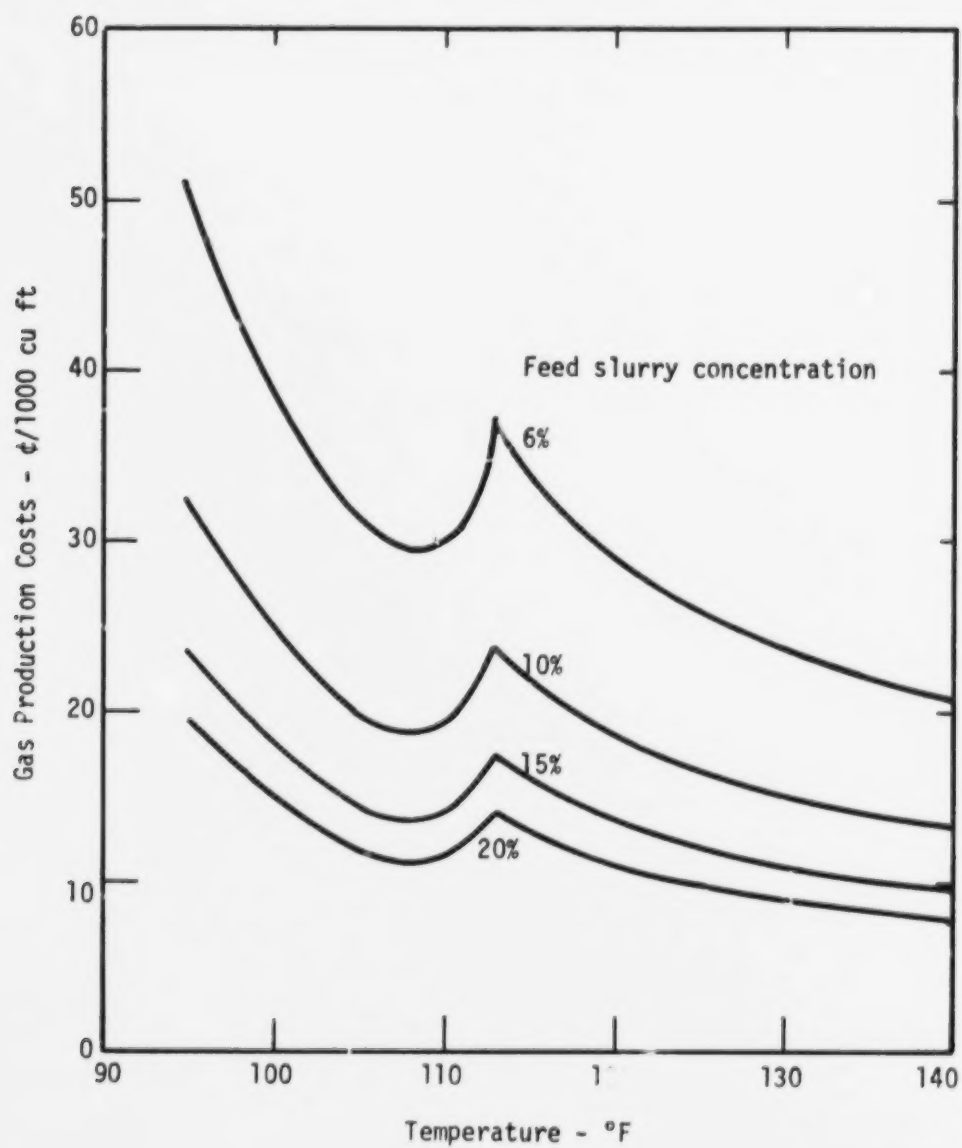


Figure 27. Gas Production Costs for Various Temperatures and Feed Slurry Concentrations

Operation at the apparent optimum conditions will produce a gas at a cost of eight to ten cents per 1000 cu ft. If this gas is marketed at \$0.75 per million BTU, a return of \$2.50 to \$3.00 per ton of dry refuse can be expected. If a shredding and separation cost of \$2.00 per ton is deducted, there is still a revenue of \$0.50 to \$1.00 per ton of dry refuse received. On the basis of the refuse characteristics given in Table 28, the revenue would be \$0.36 to \$0.72 per ton of refuse as received. Therefore, by intercepting the refuse on its way to the disposal site and recovering the gas, an income of \$0.36 to \$0.72 per ton can be recognized.

The above benefits have not included benefits relating to land disposal of the refuse. Processing for methane recovery reduces the quantity of solids to be disposed. The inert fraction plus the dewatered reactor residue yield about 80 tons of residue per 100 tons of refuse as received. However, this residue is more dense and will require only about one-third the volume of landfill required for unprocessed refuse. There will be definite cost reductions both for transportation to the disposal site as well as disposal costs. Because these costs are highly dependent upon local conditions, they were not included in this cost analysis.

An additional cost factor not included is the cost for the disposal of the sludge from the water pollution control plant. These costs may range from as low as \$15 per ton of dry solids to as high as \$60 per ton of dry solids. Again these costs are highly dependent upon local conditions. A system processing 100 tons per day of refuse (as received) would also process 2.9 tons of dry sludge solids from the water pollution control plant. The cost credit for processing this sludge could range from \$43.50 to \$174 per 100 tons of refuse processed. This would add a \$0.435 to \$1.74 cost credit to the \$0.36 to \$0.72 per ton revenue generated from the gas production. Therefore, the net benefits that can accrue from this system may range from \$0.80 to \$2.46 per ton of refuse processed. As stated previously, this figure does not include any allowance for reduction in disposal costs. Also, no credit has been claimed for any salvage value of the metals that may be separated from the refuse.

REFERENCES

- Andrews, J. F., Cole, R. D., and Pearson, E. A., "Kinetics and Characteristics of Multistage Methane Fermentations," Univ. of California, Berkeley, SERL Report No. 64011, December 1964.
- Buswell, A. M., and Mueller, H. F., "Mechanisms of Methane Fermentation," *Industrial and Engineering Chemistry*, 44, 550-552, 1952.
- Chan, D. B., and Pearson, E. A., "Comprehensive Studies of Solid Wastes Management-Hydrolysis Rate of Cellulose in Anaerobic Fermentation," Univ. of California, Berkeley, SERL Report No. 70-3, October 1970.
- Chemical and Engineering News*, 48, 34-35, March 2, 1970(a).
- Chemical and Engineering News*, 48, 8, March 30, 1970(b).
- Chemical and Engineering News*, 48, 11, April 13, 1970(c).
- Drobny, N. L., Hull, H. E., and Testin, R. F., "Recovery and Utilization of Municipal Solid Waste: A Summary of Available Cost and Performance Characteristics of Unit Processes and Systems," Public Health Service Publication No. 1908, Washington, D. C., 1971.
- Eckenfelder, W. W., *Industrial Water Pollution Control*, McGraw-Hill, New York, 1966.
- Eckenfelder, W. W., and Adams, C. A., Jr., "Design and Economics of Joint Wastewater Treatment," *Jour. of the San. Engineering Division*, ASCE, 98, SA 1, 153, February 1972.
- Dubos, R. J., "Influence of Environmental Conditions on the Activities of Cellulose Decomposition Bacteria in the Soil," *Ecology*, 9, 12, 1928.
- Eliassen, R., "Solid Waste Management - A Comprehensive Assessment of Solid Waste Problems, Practices, and Needs," Office of Science and Technology, Executive Office of the President, Washington, D. C., May 1969.
- Enebo, L., "Formation of Reducing Sugars in Thermophilic Cellulose Fermentation," *Acta Chem. Scand.*, 3, 975, 1949.

- Enebo, L., and Pehrson, S., "Thermophilic Digestion of Domestic Sewage Sludge and Cellulose Materials," *Acta Polytech. Scand.*, 281, 36, 1960.
- Engineering News Record*, 189, 25, 96, June 22, 1972.
- Gaudy, A. F., Jr., "Colorimetric Determination of Protein and Carbohydrate," *Industrial Water and Wastes*, 7, 17-22, January-February 1962.
- Golueke, C. G., "Temperature Effects on Anaerobic Digestion of Raw Sewage Sludge," *Sewage and Industrial Wastes*, 30, 1225, 1958.
- Golueke, C. G., and McGauhey, P. H., "Comprehensive Studies of Solid Wastes Management. First Annual Report," Univ. of California, Berkeley, SERL Report No. 67-7, May 1967.
- Golueke, C. G., "Comprehensive Studies of Solid Wastes Management. Third Annual Report," Univ. of California, Berkeley, SERL Report No. 70-2, June 1970.
- Ham, R. K., Porter, W. K., and Reinhardt, J. J., "Refuse Milling for Landfill Disposal," *Public Works*, December 1971.
- Heukelekian, H., "Decomposition of Cellulose in Fresh Sewage Solids," *Industrial and Engineering Chemistry*, 19, 928-930, 1927.
- Heukelekian, H., "Studies on the Decomposition of Cellulose," *Public Health Engineering Abstract*, E-545 32, 1928.
- Hungate, R. E., "The Anaerobic Mesophilic Cellulolytic Bacteria," *Bact. Rev.*, 14, 1-49, 1950.
- Koehler, L. H., "Differentiation of Carbohydrates by Anthrone Reaction Rate and Color Intensity," *Analytical Chemistry*, 24, 1576-1579, October 1952.
- Lawrence, A. W., and McCarty, P. L., "Kinetics of Methane Fermentation in Anaerobic Waste Treatment," Dept. of Civil Engineering, Stanford Univ., 1967.
- Maki, L. R., "Experiments on the Microbiology of Cellulose Decomposition in a Municipal Sewage Plant," *Antonie Van Leeuwenhoek*, 20, 185-200, 1954.
- McBee, R. H., "The Culture and Physiology of a Thermophilic Cellulose-Fermenting Bacterium," *Journal of Bacteriology*, 56, 653, 1948.

- McCarty, P. L., "Anaerobic Waste Treatment Fundamentals - Part One, Chemistry and Microbiology," *Public Works*, 95, 107-112, September 1964(a).
- McCarty, P. L., "Anaerobic Waste Treatment Fundamentals - Part Two, Environmental Requirements and Control," *Public Works*, 95, 123-126, October 1964(b).
- McCarty, P. L., "Anaerobic Waste Treatment Fundamentals - Part Three, Toxic Materials and Their Control," *Public Works*, 95, 91-94, November 1964(c).
- McCarty, P. L., "Anaerobic Waste Treatment Fundamentals - Part Four, Process Design," *Public Works*, 95, 95-99, December 1964(d).
- McKinney, R. E., and Conway, R. A., "Chemical Oxygen in Biological Waste Treatment," *Sewage and Industrial Wastes*, 29, 1097-1106, October 1957.
- McMahon, T. E., "Design of a Mechanically Mixed Digester," *The Bulletin of Engineering and Architecture*, No. 54, Univ. of Kansas, Lawrence, 1965.
- Municipal Refuse Disposal*, 2nd Ed., American Public Works Association, Chicago, Illinois, 1966.
- Pfeffer, J. T., "Increased Loadings on Digesters with Recycle of Digested Solids," *Journal of Water Pollution Control Federation*, 40, 1920-1933, November 1968.
- Pohland, F. G., and Bloodgood, D. E., "Laboratory Studies on Mesophilic and Thermophilic Anaerobic Sludge Digestion," *Journal of Water Pollution Control Federation*, 35, 11, 1963.
- Porter, W. K., "An Economic and Field Evaluation of Pulverizing Refuse for Landfill Disposal," The Univ. of Wisconsin, Madison, April 1970.
- Pringsheim, H., "Über den Fermentativen Abbau der Cellulose," *Zeitschrift für Physiologische Chemie*, 78, 266-291, 1912.
- Rankin, R. S., "Digester Capacity Requirements," *Sewage Works Journal*, 20, 478-489, 1948.
- Reese, E. T., Siu, R. G. H., and Levinson, H. S., "The Biological Degradation of Soluble Cellulose Derivation and Its Relationship to the Mechanism of Cellulose Hydrolysis," *Bact. Rev.*, 59, 485-497, 1950.

- Reese, E. T., and Levinson, H. S., "A Comparative Study of the Breakdown of Cellulose by Microorganisms," *Physiologia Plantarum*, 5, 345-366, 1952.
- Reese, E. T., "Enzymatic Hydrolysis of Cellulose," *Applied Microbiology*, 4, 39-45, 1956.
- Rietema, K., and Verver, C. G., [Eds.], *Cyclones in Industry*, Elsevier Publishing Co., Amsterdam, 1961.
- Snell, F. D., and Snell, C. T., *Colorimetric Methods of Analysis*, Volume 3, D. Van Nostrand, Princeton, New Jersey, 1961.
- Speece, R. E., and McCarty, P. L., "Nutrient Requirements and Biological Solids Accumulation in Anaerobic Digestion," *Advances in Water Pollution Research*, 2, 305-322, Eckenfelder, W. W. [Ed.], Pergamon Press, New York, 1964.
- Standard Methods for the Examination of Water and Wastewater*, 12th Ed., American Public Health Association, New York, 1965.
- Tappeiner, H. von, "Untersuchungen über die Gähryng der Cellulose, insbesondere über deren Lösung in Darmkanal," *Zietschrift für Biologie*, 20, 52-134, 1884.
- Water Pollution Control Federation, *Anaerobic Sludge Digestion - WOP 16*, Washington, D. C., 1966.
- Viles, F. J., Jr., and Silverman, L., "Determination of Starch and Cellulose with Anthrone," *Analytical Chemistry*, 21, 950-953, August 1949.

APPENDIX A

DETERMINATION OF CELLULOSE AND NON-CELLULOSE CARBOHYDRATES

Apparatus

Waring blender
hotplate or bunsen burner
filtration apparatus
 250 ml sidearm erlenmeyer flasks fitted to vacuum system
 25 ml Gooch crucibles with asbestos pads
 32 mm filtration holders for Gooch crucibles
Spectronic 20 set at 620 mμ
1 cm tubes for Spectronic 20
volumetric pipets: 1 ml, 2 ml, 4 ml
graduated pipet: 1 ml in 1/10
ice bath
stirring rods

Reagents

60 percent (by volume) H_2SO_4 - cool
anthrone reagent = 1 g anthrone/ 1 conc. H_2SO_4 must be between
 four hours and nine days old
dextrose solution - 100 mg anhydrous dextrose/ 1 H_2O

Procedure

Preparation of Gooch crucibles with mats

Stir about 10 g asbestos into 1 l H_2O

Pour this mixture through the crucibles until a mat about
1/8 in. thick is formed. Make a new asbestos mixture when
the pads start to be made of coarse fibers.

Wash the pads with 60 percent H_2SO_4 and then with H_2O

Dry overnight in a 70° oven and store in a dessicator

Standards

Standards must be prepared at the time the samples are
tested because the anthrone reagent is unstable.

Put H_2O and dextrose solution into 11 colorimeter tubes in
the amounts indicated:

Tube	H ₂ O ml	Dextrose ml
1	2	0
2	1.8	.2
3	1.6	.4
4	1.4	.6
5	1.2	.8
6	1.0	1.0
7	.8	1.2
8	.6	1.4
9	.4	1.6
10	.2	1.8
11	0	2

Separate the tubes before adding anthrone, because the color development is very sensitive to heat.

Add 4 ml anthrone reagent rapidly to each tube. If a safety pipette filter is used, use the smaller bulb to expel the reagent. Stir well immediately after addition. Be sure that there is no yellower region at the bottom of the tube.

Allow each tube to air cool for 10 minutes, then cool in an ice bath.

Wipe each tube carefully to remove water drops and fingerprints.

Do not put ice-cold tubes into the colorimeter. They will fog up.

Let the colorimeter warm up for about 20 minutes before standardizing.

Tube 1 is the reference blank. Use it to set 100 percent transmittance on the colorimeter. Restandardize the colorimeter during the test.

Determine the transmittance of each standard tube and plot transmittance against ug carbohydrate on semilog graph paper.

Samples

It is advisable to do this test in duplicate, because variation is easily introduced into the procedure.

Dilute the sample in H_2O until a 2 ml aliquot contains between 10 and 200 ug carbohydrates. If the undiluted sample is very thick or heterogeneous, use a blender to mix the first dilution.

Boil 50 ml of the diluted sample for 15 minutes.

Allow to air cool, then bring back to its original 50 ml volume with H_2O .

Filter this solution through the Gooch crucibles into erlenmeyer flasks. The filtrate will contain all non-cellulose carbohydrates which have been broken down in boiling. The residue will contain cellulose. Set aside this first filtrate.

Use clean erlenmeyer flasks for the second filtration. Add about 20 ml of exactly 50 ml of 60 percent H_2SO_4 to the crucibles. With a stirring rod, gently scrape the residue from the surface of the asbestos pad, and mix the residue into the acid.

Allow the mixture to digest 15 to 30 minutes, then draw the acid in the crucible and the balance of the 50 ml into the flask.

Put a 2 ml aliquot of the first filtrate in a colorimeter tube. Follow the procedure given for the standards for the addition of anthrone, mixing and cooling. Determine ug non-cellulose carbohydrate in the 2 ml aliquot from the colorimeter and the standard transmission line.

Put 2 ml H_2O and 0.5 ml of the second filtrate into a colorimeter tube. Mix by gentle shaking and allow it to air cool. Follow the procedure given for the standards for the addition of anthrone, mixing and cooling. Determine ug cellulose present in the 0.5 ml aliquot from the colorimeter reading and the standard transmission line.

APPENDIX B

FILTER TEST LEAF PROCEDURE

1. Determine the percent solids in the slurry before starting test work and record.
2. With the test leaf tubing valued off, read the static vacuum pressure and record in inches mercury.
3. Place two liters or more of sludge in an open beaker big enough to accommodate the test leaf.
4. Immerse the fabric-covered test leaf in the agitated slurry (about two inches below surface). Apply full vacuum at time zero and filter for the predetermined form time. Record form time and vacuum.
5. Remove the leaf from the slurry after the form portion of the cycle is completed and allow air to be pulled through the cake and leaf for the remainder of the cycle (dry time). Record dry time and vacuum reading.
6. At the end of the drying cycle shut vacuum off and allow tubing to drain completely. Record the amount of filtrate obtained.
7. Place the filter cake on a dry tared container, weigh and record.
8. Dry the cake in an oven at 103°C, reweigh and record cake dry weight.
9. Calculate filter yield in pounds per hour per square foot of area, and percent moisture in the cake formed.
10. Determine the suspended solids concentration of the filtrate.

Appendix C

Table 35. Alkalinity and Volatile Acids Determinations - 35°C

	Alkalinity - mg/ℓ Ca CO ₃								Volatile Acids - mg/ℓ Acetic							
	1	2	3	4	5	6	7	8	1	2	3	4	5	6	7	8
1530		1620	1820	1840	1910	2280	2520	2850	424	408	408	440	344	188	282	376
1550		1550	1510	1935	-	2360	2590	3110	484	390	-	563	148	465	422	267
1150		1215	1410	1730	1910	2250	2680	2920	644	358	210	160	232	359	865	179
1280		1280	1665	1935	1935	2000	2500	3130	292	586	293	464	171	220	293	263
1350		1325	1640	1960	2040	2360	2630	3280	256	281	159	122	134	159	146	208
1240		1200	1510	1910	2040	2500	2700	3130	244	281	195	171	244	232	220	98
1260		1260	1570	1890	2050	2280	2650	3240	670	364	182	195	183	159	354	278
1240		1380	1560	1960	2040	2808	2640	3370	646	525	232	183	281	134	281	390
-		1305	1530	1945	2200	2680	2740	3400	557	452	183	183	242	183	183	281
1120		1210	1530	1870	2140	2510	2790	3510	480	420	288	264	262	240	312	300
1305		1260	1485	1775	1910	2420	2660	3310	414	414	260	331	442	262	331	497
1150		1215	1418	1730	1890	2360	2650	3260	454	406	615	522	360	278	349	487
1215		1190	1370	1690	1935	2320	2658	3350	418	372	418	248	395	371	278	255
1190		1280	1440	1755	1960	2520	2725	3370	558	337	325	290	313	372	418	418
1190		1240	1350	1685	2070	2340	2810	3150	475	360	232	313	349	360	545	626
1260		1460	1530	1800	2140	2740	2930	3330	470	543	399	554	318	338	294	410
1350		1510	1485	1850	2040	2540	2810	3500	380	299	342	348	205	353	174	353
1395		1485	1530	1820	2050	2480	2810	3370	327	405	250	260	200	211	264	211
1395		1395	1530	1755	2040	2410	2790	3460								
1350		1390	1415	1708	2060	2340	2790	3440								
1285		1420	1400	1760	2050	2340	2320	3420								
1350		1485	1575	1800	2120	2380	2820	3550								
1260		1305	1370	1820	2070	2290	2790	3440								

Table 36. Alkalinity and Volatile Acids Determinations - 40°C

Alkalinity - mg/L CaCO ₃										Volatile Acids - mg/L Acetic							
1	2	3	4	5	6	7	8	1	2	3	4	5	6	7	8		
1190	1170	1418	1556	1578	1910	1665	2115	240	278	192	229	426	655	1390	1115		
1260	1260	1690	1730	1930	2340	2540	2790	188	200	177	216	497	730	906	911		
1239	1282	1640	1775	2000	2205	1757	2930	249	177	194	216	475	610	825	1355		
1330	1330	1710	1780	1980	2480	2740	2970	505	293	298	555	430	650	550	224		
1210	1350	1330	1780	2020	2450	2860	3180	170	293	218	205	218	580	650	975		
1200	1315	1700	1790	1930	2290	2860	3270	565	550	640	465	336	585	665	900		
1310	1260	1465	1580	1730	1980	2070	2050										
1215	1370	1580	1670	1730	2160	2140	2140										

Table 37. Alkalinity and Volatile Acids Determinations - 50°C

Alkalinity - mg/L CaCO ₃										Volatile Acids - mg/L Acetic							
1	2	3	4	5	6	7	8	1	2	3	4	5	6	7	8		
1360	1280	1620	1730	1560	1760	1930	2040	-	150	59	288	133	239	199	218		
1080	1060	1040	1320	1330	1760	2280	2560	-	271	204	229	254	350	330	482		
1170	1180	1090	1260	1620	1710	2140	2430	292	210	296	256	267	388	378	505		
878	968	945	1120	1500	1570	1980	2340										
1290	1280	1560	1540	1710	2090	-	-										

Table 38. Alkalinity and Volatile Acids Determinations - 55°C

1	Alkalinity - mg/ℓ CaCO ₃								Volatile Acids - mg/ℓ Acetic							
	2	3	4	5	6	7	8		1	2	3	4	5	6	7	8
990	1376	1530	1530	1530	1580	1510	1580		102	39	67	70	146	210	350	552
1120	1460	1600	1690	1390	1550	1640	-		92	69	71	92	262	290	360	339
1080	1330	1540	1550	1460	1430	1430	1920		143	65	-	58	216	164	236	223
1200	1400	1550	1640	1560	1520	1440	1840		-	-	-	99	-	198	237	216
890	1160	1220	1510	1560	1730	1550	2000									
960	1120	1150	1360	1340	1610	1500	2000									
1120	1250	1180	1340	1210	1390	1380	1970									

APPENDIX D

PROGRAM INPUT DATA

MIXING POWER REQUIRED HP/1000 CU FT AT 4 PERCENT CONCENTT. = 0.185
POWER COST CENTS/KWH = 1
EXTERNAL TEMPERATURE DEGREES F. = 45
INLET SLUDGE TEMPERATURE DEGREES F. = 35
HEATING COST CENTS/1000 CU FT = 55
SLUDGE DEWATERING POWER REQUIRED HP/SQ. FT OF FILTER AREA = 0.125
GAS VALUE CENTS/1000 CU FT = 45
PLANT CAPACITY TONS PER DAY = 100.0

```

1  DIMENSION IEM(10),UAY(6),VNL(6),ACC(6),FCC(6)
2  DIMENSION GP(6,10),VSD(6,10),SC(6,10),HL(6,10),SH(6,10),MV(6,10)
3  DIMENSION TH(6,10),HC(6,10),FY(6,10),FA(6,10),DC(6,10),TC(6,10)
4  DIMENSION GF(6,10),NR(6,10),MP(6,10)
5  DIMENSION FJ(7)
6  REAL M,NH,MP
7  FJ(1)=3.0
8  FJ(2)=6.0
9  FJ(3)=10.0
10 FJ(4)=15.0
11 FJ(5)=20.0
12 FJ(6)=25.0
13 FJ(7)=30.0
14 P1=198.0
15 P2=8.06
16 P3=0.45
17 T0=45.0
18 T1=35.0
19 M=2.0E5
20 READ(5,100)(IEM(I),IT=1,10)
21 READ(5,101)(UAY(ID),ID=1,6)
22 READ(5,102)((GP(ID,IT),ID=1,6),IT=1,10)
23 READ(5,102)((VSD(ID,IT),ID=1,6),IT=1,10)
24 WRITE(6,307)
25 WRITE(6,201)
26 WRITE(6,202)
27 WRITE(6,312)(IEM(I1),(GP(ID,IT),ID=1,6),IT=1,10)
28 WRITE(6,308)
29 WRITE(6,201)
30 WRITE(6,202)
31 WRITE(6,312)(TFM(IT),(VSD(ID,IT),ID=1,6),IT=1,10)
32 UN 60 J=1,7

```

```

33      UN 40 ID=1,6
34      FC(ID)=FJ(J)
35      UN 50 IT=1,10
36      SC(ID,IT)=(1.-0.00/30*VSU(ID,IT))+FC(ID)
37
38      50 CONTINUE
39      40 CONTINUE
40      UN 20 ID=1,6
41      T=DAY(ID)
42      VOL(ID)=M*1/FC(ID)/62.5*1.0E2
43      V=VOL(ID)
44      V1=V*1.0E-3
45      ACC(ID)=0.0782*(1.65*V1+17.*V1**0.13)*1.0E-3
46      UN 30 IT=1,10
47      MP(ID,IT)=SC(ID,IT)*11.95*V/4.0*1.0E-9
48      T2=TEM(IT)
49      HL(ID,IT)=36.*V**0.67*(T2-T0)*1.0E-6
50      T=DAY(ID)
51      SH(ID,IT)=V*62.5*(T2-T1)/T*0.25*1.0E-6
52      G=GP(ID,IT)
53      MV(ID,IT)=((1.48+0.16*(T2-35.0))*M*1.0E-6+G)
54      TH(ID,IT)=HL(IT)+SH(ID,IT)+MV(ID,IT)
55      HC(ID,IT)=P1*TH(ID,IT)*1.0E-6
56      SC1=SC(ID,IT)
57      IF(SC1.LI.1.8) FY1=1.
58      IF(SC1.GE.1.8) FY1=2.34*SC1-3.2
59      FY(ID,IT)=FY1
60      FA(ID,IT)=2.6*V/T*SC(ID,IT)*1.0E-2/FY1
61      DC(ID,IT)=P2*FA(ID,IT)*1.0E-6
62      TC1=MP(ID,IT)+FC(ID,IT)+DC(ID,IT)+ACC(ID)
63      TC(ID,IT)=TC1
64      GF(ID,IT)=TC1/A/G*1.0E11/360.
65      NB(ID,IT)=P3*M*G*360.*1.0E-9-TC1

```

65	30 CONTINUE	
66	20 CONTINUE	
67	WRITE(6,204)	
68	WRITE(6,600) FJ(J)	
69	WRITE(6,201) (DAY(ID),ID=1,6)	
70	WRITE(6,200) (C(ID),ID=1,6)	
71	WRITE(6,205)	
72	WRITE(6,600) FJ(J)	
73	WRITE(6,201) (DAY(ID),ID=1,6)	
74	WRITE(6,202)	
75	WRITE(6,203) (TFM(IT),(SC(ID,IT),ID=1,6),IT=1,10)	
76	WRITE(6,206)	
77	WRITE(6,600) FJ(J)	
78	WRITE(6,201) (DAY(ID),ID=1,6)	
79	WRITE(6,200) (VCL(ID),ID=1,6)	
80	WRITE(6,207)	
81	WRITE(6,600) FJ(J)	
82	WRITE(6,201) (DAY(ID),ID=1,6)	
83	WRITE(6,280) (ACC(ID),ID=1,6)	
84	WRITE(6,208)	
85	WRITE(6,600) FJ(J)	
86	WRITE(6,201) (DAY(ID),ID=1,6)	
87	WRITE(6,202)	
88	WRITE(6,203) (TEM(IT),(MP(ID,IT),ID=1,6),IT=1,10)	
89	WRITE(6,209)	
90	WRITE(6,600) FJ(J)	
91	WRITE(6,201) (DAY(ID),ID=1,6)	
92	WRITE(6,202)	
93	WRITE(6,203) (TEM(IT),(PL(ID,IT),ID=1,6),IT=1,10)	
94	WRITE(6,210)	
95	WRITE(6,600) FJ(J)	
96	WRITE(6,201) (DAY(ID),ID=1,6)	


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97 WRITE(6,202)
98 WRITE(6,203) (TEM(IT),(SM(ID,IT),ID=1,6),IT=1,10)
99 WRITE(6,211)
100 WRITE(6,600) FJ(J)
101 WRITE(6,201) (DAY(IU),IU=1,6)
102 WRITE(6,202)
103 WRITE(6,203) (TEM(IT),(NV(ID,IT),ID=1,6),IT=1,10)
104 WRITE(6,212)
105 WRITE(6,600) FJ(J)
106 WRITE(6,201) (DAY(IU),IU=1,6)
107 WRITE(6,202)
108 WRITE(6,203) (TEM(IT),(IW(ID,IT),ID=1,6),IT=1,10)
109 WRITE(6,213)
110 WRITE(6,600) FJ(J)
111 WRITE(6,201) (DAY(IU),IU=1,6)
112 WRITE(6,202)
113 WRITE(6,203) (TEM(IT),(HC(ID,IT),ID=1,6),IT=1,10)
114 WRITE(6,214)
115 WRITE(6,600) FJ(J)
116 WRITE(6,201) (DAY(IU),IU=1,6)
117 WRITE(6,202)
118 WRITE(6,203) (TEM(IT),(FY(ID,IT),ID=1,6),IT=1,10)
119 WRITE(6,215)
120 WRITE(6,600) FJ(J)
121 WRITE(6,201) (DAY(IU),IU=1,6)
122 WRITE(6,202)
123 WRITE(6,203) (TEM(IT),(FA(ID,IT),ID=1,6),IT=1,10)
124 WRITE(6,216)
125 WRITE(6,600) FJ(J)
126 WRITE(6,201) (DAY(IU),IU=1,6)
127 WRITE(6,202)
128 WRITE(6,203) (TEM(IT),(DC(ID,IT),ID=1,6),IT=1,10)

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129 WRITE(6,217)
130 WRITE(6,600) FJ(J)
131 WRITE(6,201) (PAY(10),10=1,6)
132 WRITE(6,202)
133 WRITE(6,203) (TEM(11),(1C(ID,11),ID=1,6),IT=1,10)
134 WRITE(6,200)
135 WRITE(6,600) FJ(J)
136 WRITE(6,201) (PAY(10),10=1,6)
137 WRITE(6,202)
138 WRITE(6,203) (TEM(11),(GF(ID,11),ID=1,6),IT=1,10)
139 WRITE(6,300)
140 WRITE(6,600) FJ(J)
141 WRITE(6,201) (PAY(10),10=1,6)
142 WRITE(6,202)
143 WRITE(6,203) (TEM(11),(NR(ID,11),ID=1,6),IT=1,10)
144 CONTINUE
145 60 100 FORMAT(10X,10F7.2)
146 101 FORMAT(10X,6F4.0)
147 102 FORMAT(10X,6F7.3)
148 200 FORMAT(1H1,///,49X,28HGAS COST ,CENTS PER 1000 CF ,/// )
149 201 FORMAT(1H ,18X,21HRETENTION TIME (DAYS),6(F10.1,3X),///)
150 202 FORMAT(1H ,19X,14HTEMP (DEG. F) )
151 203 FORMAT(1H ,21X,F6.1,12X,E10.3,3X,E10.3,3X,E10.3,3X,
1  ,E10.3,3X,E10.3,3X /)
152 204 FORMAT(1H1,///,40X,33HFEEED CONCENTRATION ( PERCENT) ///)
153 205 FORMAT(1H1,///,40X,30HSLURRY CONTENT ( PERCENT ) ///)
154 206 FORMAT(1H1,///,40X,29HDIGESTER VOLUME ( CU FT ) ///)
155 207 FORMAT(1H1,///,40X,45HAMORTIZED CAPITAL COST ( MILL.$ PER YEAR )
1 ///)
156 208 FORMAT(1H1,///,40X,36HMIXING POWER (MILL. $ PER YEAR ) ///)
157 209 FORMAT(1H1,///,40X,10HHEAT LOSS ,
1 23H(MILL. BTU PER DAY ) ///)

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158 210      FORMAT(1H1,/,/,40X,15HSLUDGE HEATING ,
159      1 23H(MILL. BTU PER DAY ) ///)
160 211      FORMAT(1H1,/,/,40X,12HWATER VAPOR ,
161      1 23H(MILL. BTU PER DAY ) ///)
162 212      FORMAT(1H1,/,/,40X,43HTOTAL HEAT REQUIRED (MILLC BTU -&R DAO= //
163      1)
164 213      FORMAT(1H1,/,/,40X, 35HHEATING COST (MILL. $ PER YEAR ) ///)
165 214      FORMAT(1H1,/,/,40X,46HFILTER YIELD (LB PER SQUARE FEET PER HOUR
166      1) ///)
167 215      FORMAT(1H1,/,/,40X,39HREQUIRED FILTER AREA (SQUARE FEET ) ///)
168 216      FORMAT(1H1,/,/,40X,39HDEWATERING COST (MILL. $ PER YEAR ) ///)
169 217      FORMAT(1H1,/,/,40X,34HTOTAL COST (MILL. $ PER YEAR ) ///)
170 280      FORMAT(1H ,41X,F10.4,3X,E10.4,3X,F10.4,3X,E10.4,3X
171      1 ,E10.4,3X //)
172 300      FORMAT(1H1,/,/,49X,32HNET RENFFIT IN MILL. $ PER YEAR // )
173 302      FORMAT(1H ,19X,14HTEMP (DEG. F) ,10F6.1 //)
174 307      FORMAT(1H1,/,/,20X,47HTABLE 7 GAS PRODUCTIONB (CU FT / LB SOLID
175      1) )
176 308      FORMAT(1H1,/,/,20X,49HTABLE 8 VOLATILE SOLID DESTRUCTION ( PERCE
177      :NT ) )
178 312      FORMAT(1H ,21X,F6.1,12X,6F7.3)
179 600      FORMAT(1H ,32X, 20HFEED CONCENTRATION = ,F6.1,3X,11H (PERCENT)
180      1 ///)
181      WRITE(6,400)
182      400 FORMAT(1H1)
183      STOP
184      END
185      **WARNING** FORMAT STATEMENT 302 IS UNREFERENCED

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SENTRY

Appendix E

Table 39. Reactor Slurry Concentration Under Various Operating Conditions - Percent Solids

Temperature °F	Reactor Retention Time - Days					
	4	8	10	15	20	30
Feed Slurry Concentration - 6% Solids						
95.0	5.24	4.75	4.59	4.40	4.39	4.34
104.0	4.73	4.38	4.26	4.04	3.90	3.77
107.8	4.59	4.29	4.18	3.96	3.85	3.71
109.4	4.59	4.31	4.22	4.00	3.92	3.77
113.0	4.87	4.61	4.53	4.34	4.29	4.17
118.4	4.56	4.29	4.18	4.06	3.92	3.78
127.4	4.18	3.88	3.78	3.66	3.54	3.44
132.8	3.96	3.71	3.63	3.53	3.41	3.34
140.0	3.69	3.55	3.51	3.39	3.31	3.29

Feed Slurry Concentration - 10% Solids						
95.0	8.73	7.91	7.66	7.33	7.32	7.23
104.0	7.88	7.29	7.09	6.73	6.50	6.28
107.8	7.65	7.15	6.97	6.60	6.41	6.18
109.4	7.66	7.19	7.04	6.66	6.53	6.28
113.0	8.11	7.69	7.55	7.23	7.15	6.95
118.4	7.60	7.15	6.97	6.76	6.53	6.31
127.4	6.97	6.47	6.30	6.10	5.90	5.73
132.8	6.60	6.18	6.04	5.88	5.69	5.56
140.0	6.15	5.92	5.85	5.65	5.52	5.49

Feed Slurry Concentration - 15% Solids						
95.0	13.1	11.9	11.5	11.0	11.0	10.8
104.0	11.8	10.9	10.6	10.1	9.8	9.4
107.8	11.5	10.7	10.5	9.9	9.6	9.3
109.4	11.5	10.8	10.6	10.0	9.8	9.4
113.0	12.2	11.5	11.3	10.8	10.7	10.4
118.4	11.4	10.7	10.5	10.1	9.8	9.5
127.4	10.5	9.7	9.5	9.2	8.9	8.6
132.8	9.9	9.3	9.1	8.8	8.5	8.3
140.0	9.2	8.9	8.8	8.5	8.3	8.2

Table 39. Continued

Temperature °F	Reactor Retention Time - Days					
	4	8	10	15	20	30
Feed Slurry Concentration - 20% Solids						
95.0	17.5	15.8	15.3	14.7	14.6	14.5
104.0	15.8	14.6	14.2	13.5	13.0	12.6
107.8	15.3	14.3	13.9	13.2	12.8	12.4
109.4	15.3	14.4	14.1	13.3	13.1	12.6
113.0	16.2	15.4	15.1	14.5	14.3	13.9
118.4	15.2	14.3	13.9	13.5	13.1	12.6
127.4	13.9	12.9	12.6	12.2	11.8	11.5
132.8	13.2	12.4	12.1	11.8	11.4	11.1
140.0	12.3	11.8	11.7	11.3	11.0	11.0

Table 40. Mixing Costs for Various Temperatures and Retention Times - \$1000 per Year

Temperature °F	Reactor Retention Time - Days					
	4	8	10	15	20	30
95.0	3.34	6.05	7.32	10.5	14.0	20.7
104.0	3.01	5.58	6.78	9.7	12.4	18.0
107.8	2.93	5.47	6.66	9.5	12.3	17.7
109.4	2.93	5.50	6.73	9.6	12.5	18.0
113.0	3.10	5.88	7.22	10.4	13.7	19.9
118.4	2.91	5.47	6.66	9.7	12.5	18.1
127.4	2.67	4.95	6.02	8.8	11.3	16.4
132.8	2.52	4.73	5.78	8.4	10.9	16.0
140.0	2.35	4.53	5.59	8.1	10.6	15.7

Table 41. Heating Costs for Various Operating Conditions - \$1000 per Year

Temperature °F	Reactor Retention Time - Days					
	4	8	10	15	20	30
Feed Slurry Concentration - 6% Solids						
95.0	11.8	13.0	13.5	14.4	15.1	16.4
104.0	14.1	15.4	15.9	17.0	17.9	19.5
107.8	15.0	16.3	16.8	18.0	19.0	20.6
109.4	15.4	16.6	17.2	18.4	19.3	21.0
113.0	15.8	17.1	17.7	18.9	19.9	21.7
118.4	17.3	18.7	19.3	20.6	21.7	23.7
127.4	19.6	21.2	21.9	23.3	24.6	26.7
132.8	21.0	22.7	23.4	24.9	26.2	28.4
140.0	22.9	24.6	25.3	26.9	28.3	30.6

Feed Slurry Concentration - 10% Solids						
95.0	7.5	8.5	8.8	9.5	10.0	10.9
104.0	9.1	10.1	10.5	11.3	12.0	13.1
107.8	9.7	10.7	11.1	12.0	12.7	14.0
109.4	10.0	10.9	11.3	12.3	13.0	14.2
113.0	10.2	11.2	11.6	12.5	13.2	14.5
118.4	11.2	12.3	12.8	13.7	14.6	16.0
127.4	12.8	14.1	14.6	15.7	16.6	18.2
132.8	13.9	15.2	15.7	16.8	17.8	19.4
140.0	15.2	16.5	17.0	18.2	19.2	20.9

Feed Slurry Concentration - 15% Solids						
95.0	5.3	6.1	6.4	7.0	7.4	8.1
104.0	6.6	7.4	7.7	8.4	9.0	9.9
107.8	7.1	7.9	8.2	8.9	9.5	10.5
109.4	7.2	8.0	8.4	9.1	9.7	10.6
113.0	7.3	8.1	8.4	9.2	9.8	10.8
118.4	8.1	9.0	9.4	10.2	10.8	12.0
127.4	9.4	10.5	10.9	11.7	12.5	13.7
132.8	10.3	11.3	11.7	12.6	13.4	14.6
140.0	11.4	12.3	12.8	13.7	14.5	15.8

Table 41. Continued

Temperature	Reactor Retention Time - Days					
°F	4	8	10	15	20	30
Feed Slurry Concentration - 20% Solids						
95.0	4.2	4.9	5.2	5.6	6.0	6.6
104.0	5.3	6.0	6.3	6.9	7.4	8.1
107.8	5.7	6.4	6.7	7.4	7.8	8.7
109.4	5.8	6.5	6.8	7.5	7.9	8.8
113.0	5.8	6.6	6.8	7.5	8.0	8.8
118.4	6.6	7.4	7.7	8.3	8.9	9.9
127.4	7.7	8.6	9.0	9.7	10.3	11.4
132.8	8.4	9.3	9.7	10.4	11.1	12.2
140.0	9.4	10.2	10.6	11.4	12.1	13.1

Table 42. Dewatering Cost for a Feed Slurry Concentration of 20 Percent Solids - \$1000/Year

Temperature	Reactor Retention Time - Days					
°F	4	8	10	15	20	30
95.0	1.55	1.57	1.57	1.58	1.58	1.58
104.0	1.57	1.58	1.59	1.59	1.60	1.61
107.8	1.57	1.58	1.59	1.60	1.60	1.61
109.4	1.57	1.58	1.59	1.60	1.60	1.61
113.0	1.56	1.57	1.58	1.58	1.58	1.59
118.4	1.57	1.58	1.59	1.59	1.60	1.61
127.4	1.59	1.60	1.61	1.61	1.62	1.63
132.8	1.60	1.61	1.62	1.62	1.63	1.63
140.0	1.61	1.62	1.62	1.63	1.64	1.64

Table 43. Total Production Costs for All Operating Conditions - \$1000/Year

Temperature °F	Reactor Retention Time - Days					
	4	8	10	15	20	30
Feed Slurry Concentration - 6% Solids						
95.0	51.8	83.8	99.4	138	177	254
104.0	54.0	85.8	102	140	179	255
107.8	54.9	86.7	102	141	180	256
109.4	55.3	87.1	103	142	180	256
113.0	55.8	87.8	104	143	182	259
118.4	57.2	89.2	105	144	182	259
127.4	59.5	91.5	107	146	185	261
132.8	61.0	93.0	109	147	186	262
140.0	63.0	94.9	111	149	188	264

Feed Slurry Concentration - 10% Solids						
95.0	33.2	53.7	63.7	88.4	113	162
104.0	34.6	54.9	64.9	89.5	114	162
107.8	35.2	55.5	65.5	90.0	114	162
109.4	35.4	55.7	65.7	90.3	115	163
113.0	35.7	56.3	66.4	91.3	116	165
118.4	36.6	57.1	67.1	91.9	116	165
127.4	38.1	58.5	68.4	93.0	117	165
132.8	39.0	59.3	69.3	93.8	118	166
140.0	40.3	60.5	70.4	95.0	119	168

Feed Slurry Concentration - 15% Solids						
95.0	24.1	38.9	46.1	63.7	81.5	117
104.0	25.1	39.8	46.9	64.4	81.6	116
107.8	25.5	40.1	47.3	64.7	82.0	116
109.4	25.7	40.3	47.5	65.0	82.3	117
113.0	25.9	40.8	48.0	65.9	83.6	119
118.4	26.6	41.3	48.5	66.2	83.5	118
127.4	27.7	42.3	49.4	66.8	84.0	118
132.8	28.4	42.9	50.0	67.4	84.5	119
140.0	29.3	43.7	50.8	68.2	85.3	119

Table 43. Continued

Temperature °F	Reactor Retention Time - Days					
	4	8	10	15	20	30
Feed Slurry Concentration - 20% Solids						
95.0	19.6	31.6	37.3	51.4	65.7	93.8
104.0	20.4	32.2	37.9	51.8	65.5	92.7
107.8	20.7	32.5	38.2	52.1	65.8	92.9
109.4	20.9	32.6	38.3	52.3	66.1	93.3
113.0	21.0	33.0	38.8	53.1	67.3	95.2
118.4	21.6	33.4	39.2	53.3	67.1	94.5
127.4	22.5	34.2	39.8	53.7	67.3	94.3
132.8	23.1	34.7	40.3	54.2	67.7	94.6
140.0	23.9	35.4	41.0	54.8	68.4	95.4

Table 44. Net Benefits Accrued Under All Operating Conditions - \$1000/Year

Temperature °F	Reactor Retention Time - Days					
	4	8	10	15	20	30
Feed Slurry Concentration - 6% Solids						
95.0	-6.4	-9.2	-15.8	-45.5	-81.4	-155
104.0	21.5	10.7	2.2	-23.6	-53.9	-122
107.8	29.0	15.0	5.5	-20.0	-51.6	-120
109.4	28.3	13.0	2.9	-22.7	-56.3	-124
113.0	11.6	-5.2	-16.1	-43.8	-79.8	-150
118.4	28.4	12.6	3.0	-28.2	-58.7	-128
127.4	48.4	34.2	24.6	- 7.1	-38.4	-109
132.8	60.2	43.1	32.3	- 0.4	-32.1	-104
140.0	74.0	50.6	37.5	- 5.8	-28.2	-104

Feed Slurry Concentration - 10% Solids						
95.0	12.1	20.8	19.9	4.3	-17.6	-63.5
104.0	40.9	41.6	38.8	27.2	11.0	-29.1
107.8	48.8	46.2	42.4	31.2	13.7	-26.4
109.4	48.2	44.4	39.9	28.6	9.1	-30.5
113.0	31.7	26.3	21.1	7.5	-14.4	-56.3
118.4	48.9	44.7	40.8	23.8	7.5	-33.3
127.4	69.8	67.2	63.4	46.0	28.8	-13.2
132.8	82.2	76.7	71.7	53.3	35.8	- 8.1
140.0	96.8	85.0	77.6	60.2	40.5	- 6.9

Feed Slurry Concentration - 15% Solids						
95.0	21.2	35.6	37.5	28.9	14.1	-17.8
104.0	50.4	56.8	56.8	52.3	43.2	17.1
107.8	58.4	61.6	60.6	56.5	46.0	20.0
109.4	57.9	59.8	58.2	53.9	41.4	16.0
113.0	41.5	41.8	39.5	33.0	18.2	- 9.7
118.4	59.0	60.4	59.4	49.5	40.3	13.6
127.4	80.2	83.4	82.5	72.2	62.1	34.2
132.8	92.8	93.2	91.0	79.7	69.4	39.6
140.0	108	102	97.3	87.0	74.4	41.2

Table 44. Continued

Temperature °F	Reactor Retention Time - Days					
	4	8	10	15	20	30
	Feed Slurry Concentration - 20% Solids					
95.0	25.8	43.0	46.3	41.2	29.9	5.0
104.0	55.1	64.4	65.8	64.8	59.2	40.2
107.8	63.2	69.3	69.7	69.1	62.2	43.2
109.4	62.7	67.5	67.3	66.6	57.6	39.2
113.0	46.4	49.6	48.6	45.7	34.4	13.6
118.5	64.0	68.3	68.7	62.4	56.7	37.1
127.4	85.4	91.5	92.0	85.3	78.8	58.0
132.8	98.1	101	101	92.9	86.2	63.5
140.0	113	110	107	100	91.4	65.3

Table 45. Gas Production Costs Under Various Operating Conditions - ¢/1000 cu ft

Temperature °F	Reactor Retention Time - Days					
	4	8	10	15	20	30
Feed Slurry Concentration - 6% Solids						
95.0	51.4	50.6	53.5	67.1	83.3	116
104.0	32.2	40.0	44.1	54.1	64.4	86.3
107.8	29.5	38.4	42.7	52.4	63.1	84.6
109.4	29.8	39.1	43.8	53.6	65.5	87.1
113.0	37.2	47.8	53.3	65.0	80.3	107
118.4	30.1	39.4	43.7	56.0	66.3	88.6
127.4	24.8	32.8	36.6	47.3	56.8	77.1
132.8	22.7	30.7	34.7	45.1	54.4	74.7
140.0	20.7	29.3	33.6	43.3	52.9	74.1

Feed Slurry Concentration - 10% Solids						
95.0	33.0	32.4	34.3	42.9	53.3	73.9
104.0	20.6	25.6	28.2	34.5	41.0	54.9
107.8	18.9	24.5	27.3	33.4	40.2	53.7
109.4	19.0	25.1	28.0	34.2	41.7	55.4
113.0	23.8	30.7	34.2	41.6	51.3	68.3
118.4	19.3	25.2	28.0	35.7	42.3	56.4
127.4	15.9	20.9	23.3	30.1	36.1	48.9
132.8	14.5	19.6	22.1	28.7	34.5	47.3
140.0	13.2	18.7	21.4	27.5	33.6	46.9

Feed Slurry Concentration - 15% Solids						
95.0	24.0	23.5	24.8	31.0	38.4	53.1
104.0	15.0	18.5	20.3	24.8	29.4	39.2
107.8	13.7	17.8	19.7	24.0	28.8	38.4
109.4	13.8	18.1	20.2	24.6	29.9	39.6
113.0	17.3	22.2	24.7	30.0	37.0	49.0
118.4	14.0	18.3	20.2	25.7	30.4	40.3
127.4	11.5	15.1	16.8	21.6	25.9	34.9
132.8	10.5	14.2	16.0	20.6	24.7	33.7
140.0	9.6	13.5	15.4	19.8	24.0	33.5

Table 45. Continued

Temperature	Reactor Retention Time - Days					
°F	4	8	10	15	20	30
Feed Slurry Concentration - 20% Solids						
95.0	19.5	19.1	20.1	25.0	30.9	42.7
104.0	12.2	15.0	16.4	20.0	23.6	31.4
107.8	11.1	14.4	15.9	19.3	23.1	30.7
109.4	11.2	14.7	16.3	19.8	24.0	31.7
113.0	14.0	18.0	20.0	24.2	29.8	39.4
118.4	11.3	14.8	16.3	20.7	24.4	32.3
127.4	9.4	12.2	13.6	17.4	20.7	27.9
132.8	8.6	11.5	12.9	16.6	19.8	26.9
140.0	7.8	10.9	12.5	15.9	19.3	26.7